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PROJECT LIGNITE

QUARTERLY TECHNICAL PROGRESS REPORT NO. 5

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ANNUAL SUMMARY

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CONTENTS

I. SUMMARY OF YEAR'S DEVELOPMENTS

  A. PDU Operation
  B. PDU Design and Construction
  C. Experimentation
  D. Economic Studies

II. OBJECTIVES AND SCOPE

III. SUMMARY OF QUARTERLY PROGRESS

IV. TECHNICAL PROGRESS

  A. PDU Operation
    1. Area 01A: Coal Crushing and Grinding
    2. Area 01B: Slurry Preparation and Pumping
    3. Area 02: Slurry Preheating, Dissolving and Gas Separation
    4. Area 04: Liquid Separation and Solvent Recycle
    5. Area 05: Gas Purification and Recycle

  B. PDU Design and Construction
    1. Area 00: Buildings and Utilities
    2. Area 01B: Slurry Preparation and Pumping
    3. Area 02: Slurry Preheating, Dissolving and Pumping
    4. Area 03: Solid Liquid Separation
    5. Area 04: Liquid Separation and Solvent Recycle
    6. Area 05: Gas Purification and Recycle
    7. Area 06: SRL Hydrogenation

  C. Experimentation
    1. Direct liquefaction without solvent
    2. Reactivity of lignite after storage
    3. Phenol pretreatment tests
    4. Variability of lignite analyses
    5. University Supporting Research

  D. Economic Studies

V. CONCLUSIONS

VI. APPENDICES

  A. Laboratory Experimentation

  B. University Supporting Research

  C. Economic Studies

  D. PDU Flow Sheets

I. SUMMARY OF YEAR'S DEVELOPMENTS

A. PDU OPERATION

The continuous lignite liquefaction process development unit was operated in a shake-down mode without use of the solid-liquid separation section. Initially, a lubricating oil-fuel oil mixture was circulated and then FS120 liquefaction solvent plus nitrogen was used. The purposes of these preliminary tests were to train operators and to determine operability of equipment.

Seven runs were initiated using lignite, solvent and synthesis gas. None of these was completed to a scheduled shutdown because of difficulties with equipment and operational problems. The most serious problem encountered was the rupture of the 316 stainless-steel slurry-preheater coil. Stress corrosion cracking accelerated by the presence of a halide is believed responsible for the failure.

Operational problems with the vacuum flash system consisted primarily of plugged discharge lines and caused most of the terminations. These and other operating difficulties were corrected by modifications of process conditions or minor changes in equipment. Most operational areas performed satisfactorily with minor changes and equipment replacement. A trial yield period of 12 hours duration was made during a shakedown test to allow checking of material balance procedures and data collection methods. Additional sample points were added and a computer program written to aid in data reduction.

B. PDU DESIGN AND CONSTRUCTION

Most areas of the unit are operational with the exception of the solids-liquid separation and the SRL hydrogenation areas where construction activities are being concentrated.

In the solid-liquid separation area, the benzene pumps and instrument panels have been installed. Completion of construction is being delayed because of back ordering of junction boxes, stop-start stations and motor starters. In the SRL hydrogenation unit, the charge pump was installed and the shell for the sand bath preheater completed.

The plant control laboratory is operational with cabinets, benches, lights and electrical services installed.

Modification of equipment and improvements in instrumentation and control units are continuing as necessary to improve operability of the PDU.

C. EXPERIMENTATION

1. Tests continued in the Project Lignite Laboratory using batch-autoclave experiments to study process conditions for liquefaction of lignite. Results from these investigations are summarized as follows:

a. Using a carbon black feedstock as starting solvent, five passes of recycled solvent through successive runs indicated no significant changes in solvent quality. Liquefaction yields were nearly constant with SRL yields of 57.7 to 58.9 weight percent of MAF coal being obtained. Solvent recovery averaged slightly over 100 percent with no indications that the carbon black feedstock would not be satisfactory as initial solvent in the PDU.

b. Direct hydrogenation of lignite without solvent in hydrogen, carbon monoxide, or synthesis gas (CO & H_2) atmosphere indicated that liquefaction using pure CO without solvent gave conversions similar to these of standards tests using solvent and synthesis gas.

c. Experiments in which lignite was dried in the autoclave with solvent indicated no reduction in sulfur content or generation of solvent type material. Actually, solvent was retained by the lignite and subsequent liquefaction was reduced. Decreased liquefaction was found in all cases where dried lignite was charged.

d. Pretreatment of lignite with phenol did not improve liquefaction characteristics.

e. Storage of lignite for 70 weeks either in air, nitrogen or under water did not greatly effect liquefaction conversion. However, some decrease in filterability of reactor slurries was noted.

2. Approximately 50 experiments were completed in laboratory tests to evaluate process parameters for the conversion of solvent refined lignite (SRL) to light liquid fuels by catalytic hydrogenation in a one-liter autoclave. A few larger scale runs were made in the larger autoclaves in Project Lignite's laboratory. Observations based on the tests are summarized as follows:

a. Catalysts are required for reasonable conversions of SRL. At 450^oC, 88 percent of the SRL was converted to distillable products.

b. $Ni-Mo-Al_2O_3$ was superior to $Ni-W-SiO_2-Al_2O_3$ as a hydrogenation catalyst.

c. Catalyst conditioned in hydrogen did not change conversion of SRL but did result in increased concentration of aromatics in hydrogenation products.

d. Highest conversions of SRL to gases and distillable products were obtained with a solvent to SRL ratio of 1:1. Ratios higher or lower gave reduced conversions.

e. With reduced solvent (tetralin) concentration, concentration of aromatics in hydrogenation products increased linearly.

f. Use of FS120 as a hydrogenation solvent resulted in conversions nearly as high as with tetralin.

g. A relatively high conversion of SRL was obtained by catalytic hydrogenation without solvent.

h. Larger scale batch tests gave similar overall conversions although the yield of gaseous products increased while gasoline yield decreased.

i. Hydrogenation of SRC (from bituminous coal) under similar test conditions gave conversions essentially the same as with SRL. However, ammonia yields were 20 percent greater because of higher nitrogen content of the SRC.

3. Studies were completed on fluidization of cracking catalyst, sand, and pulverized lignite in air streams for dense-fluidization and for fast fluidization at ambient temperature. The purpose of the experiments was to establish process parameters for both modes of fluidization for possible design of a carbonization unit. The complete system consisted of a cyclone, a storage bin, a Venturi feeder, air supply flow meter, and fluidization tubes. All materials were fluidized in both modes. A summary of major observations follows:

a. In a conventional boiling bed, incipient fluidization velocities were 0.033 to 0.066 ft/sec depending on size and density of particles.

b. Static bed heights for boiling beds could be nearly doubled in height when the tube diameter was increased from 1.5 to 4 inches.

c. The lower limit of the fast fluidization regime is characterized by a critical velocity depending on particle size, shape and density and the fluidization medium.

d. A high slip velocity of particles in fast fluidization, nearly that of the superficial velocity of the medium, indicated a high degree of back mixing.

e. Typical values for fast fluidization for 100 mesh catalyst particles are air velocities of 4.33 to 7.48 ft/sec with solid feed rates to 5 lb/min giving bed densities as high as 15 lb/cu ft.

D. ECONOMIC STUDIES

In a continuing economic program for estimation of the costs of converting lignite to higher quality fuels, several studies have been made. One study of a plant for the manufacture of solvent refined lignite (SRL) included the production of the reducing gas required for liquefaction by the reforming of natural gas; this plant was modified to produce the reducing gas by the gasification of lignite in a slagging Lurgi gasifier. This modification resulted in essentially no change in the investment of about $236,712,000, but reduced the total annual operating costs after by-product credits from $77,129,000 to $54,844,000 with a resulting decrease in product cost from 83 to 65 ¢/MMBtu when evaluated by the Discounted Cash Flow (DCF) method using a discount rate of 12 percent.

The operation of the gasification plant alone to produce 121 million standard cubic feet per day of 350 Btu/CF gas was evaluated in a later study. In this case the total investment for the gasification plant was estimated at $24,984,000 and total annual operating costs after by-product credits at $5,536,000. This resulted a product selling price of 68¢/MMBtu when evaluated by the DCF method using a discount rate of 12 percent.

II. OBJECTIVE AND SCOPE

Project Lignite was established in 1972 in the Department of Chemical Engineering at the University of North Dakota for the purpose of determining the appropriate technological approach to the conversion of Northern Great Plains lignite to premium solid, liquid and gaseous fuels. The ultimate goal of the work is to make recommendations for a refinery based on lignite as feedstock, and the experimental portion of the program is intended to develop data and know-how to carry out technological steps in the refinery processes for which information for scaleup was insufficient.

In keeping with this objective, laboratory and bench-scale work have been carried out and a continuous process development unit (PDU) has been constructed, and is being operated in preliminary tests. The PDU is to demonstrate the feasibility of continuous liquefaction of lignite under pilot plant conditions. Economic studies have been carried out by an economist attached to the program, and under a project subcontract the Department of Chemistry of UND is developing information on catalytic hydrogenation of the solvent refined lignite.

The continuous process under development is the two stage conversion of lignite to fuel liquids, with solvent refined lignite as an intermediate and useful fuel product. The first stage converts 50 pounds per hour of raw lignite into solvent refined lignite by reaction with synthesis gas (carbon monoxide and hydrogen) in the presence of a hydrogen donor solvent that is continually regenerated from the lignite during processing. The mineral separations system will be a countercurrent decantation process carried out at elevated temperature and pressure. The second stage will be the catalytic hydrogenation to light-liquid-fuel blend-stocks of the solvent refined lignite produced in the first stage. Carbonization and direct hydrogenation of lignite are also to be investigated as process steps.

III. SUMMARY OF QUARTERLY PROGRESS

A. PDU OPERATION

Seven preliminary or shakedown runs were attempted with none being completed to a scheduled shut-down. Most difficulties were associated with operational problems primarily of plugged discharge lines in the vacuum flash system. These operational and minor equipment problems were correctible. However, a serious difficulty was the sudden rupture of the stainless-steel slurry-preheater coal. It appears that halide stress corrosion was responsible for the failure.

B. PDU DESIGN AND CONSTRUCTION

The PDU plant control laboratory was supplied with cabinets, benches, lights and electrical services. Considerable electrical work was completed and some equipment and instrumentation modified.

Since the most of the PDU is in operational status, major emphasis was given to the solid-liquid separation and the SRL hydrogenation areas. Solvent pumps and instrumentation panels were installed in the solid-liquid separation unit although back ordering of electrical supplies have prevented completion of electrical wiring. Progress is being made in the SRL hydrogenation area where the SRL charge pump was installed as was the shell for the sand-bath preheater.

In a shake-down run a trial yield period was established for determination of material balances and data collection procedures. Additional sample points were added to provide needed material balance data.

C. EXPERIMENTATION

In tests for direct liquefaction of lignite without solvent, use of a pure CO atmosphere as the reducing gas is more attractive than either pure hydrogen or synthesis gas (CO & H_2). Net SRL yields were similar for tests using CO without solvent and the usual tests with synthesis gas and solvent.

Results using lignite samples which had been stored for 70 weeks under a nitrogen atmosphere, in air, or under water indicated that the unconverted material was essentially the same as for the standard baseline tests. Yields of gaseous products seemed to increase while yields of SRL did not generally change.

Pretreatment of lignite with phenol prior to liquefaction had been suggested as a possible method for improving liquefaction. Analysis of data from a test series with variations in phenol pretreatment did not indicate an appreciable advantage for the pretreatment.

Batch hydrogenation tests of SRL in University supporting research indicated that hydrogen conditioning of the catalyst was not effective in increasing overall conversion and that mixing was required for high conversion even with a large excess of solvent. Tests with SRC indicated essentially the same conversions under similar test conditions as with SRL. Addition of WS_2 to the Ni-Mo-Al_2O_3 catalyst did not change conversions.

D. ECONOMICS

Continuing cost estimates culminated during this period in two studies relating to the conversion of lignite to higher quality fuels. The results of these studies are summarized in the following paragraphs.

A comparative economic evaluation was made of two kinds of plants for processing North Dakota lignite into a composite product consisting mainly of a clean burning boiler fuel called Solvent Refined Lignite (SRL) and a light liquid resembling crude oil. The first plant is a straightforward modification of the Solvent Refined Coal process (SRC) described in OCR R&D Report No. 53, Interim Report No. 1 (1970), hereinafter referred to as the Standard Solvent Refined Lignite Plant (SSRL). The second plant is also a modified version of the SRC process but with certain important alterations, the most significant of which is the substitution of a Lurgi gasification plant for the natural gas reforming plant to produce the reducing gas for liquefaction. This latter facility will be referred to henceforth as the Solvent Refined Lignite with Lurgi Plant (SRLL).

The SSRL has a processing capacity of 35M tons of lignite per stream day from which 8,850 tons of SRL and 2,830 tons of light oil are produced. The SRLL has a processing capacity of 39M tons per day from which 8,850 tons of SRL, 2,976 tons of light oil, and 3,432 tons of synthesis gas are produced.

Total investment for each plant is estimated at $237 million. Although capital costs are similar, annual operating costs show a substantial divergence with total cost being estimated at $81 million for SSRL and $57 million for SRLL before by-product credits. Evaluating product costs by the Discounted Cash Flow (DCF) method at a 12 percent discount rate yields a product selling price of 83 ¢/MM Btu for the SSRL and 65 ¢/MM Btu for the SRLL.

An economic evaluation was made of a plant designed to produce 121 million standard cubic feet per day of 350 Btu/CF gas. Major components of the system envisioned were coal preparation, gasification, purification, and sulfur recovery. The plant had a processing capacity of over 3M tons of lignite per stream day from which the low-Btu gas was produced.

Total investment was estimated at $21 million and was assumed to be 100% equity financed. Coal and labor were the major operating costs with lignite at $2.25/ton and labor at $5.00/hr. Annual operating costs were $5.6 million before by-products whose combined value was $65 thousand per year. The foregoing data were evaluated by the DCF method at a 12 percent discount rate to yield a selling price of 68¢/MM Btu for the product.

IV. TECHNICAL PROGRESS

A. PDU OPERATIONS

Seven shakedown runs were attempted. None was completed to a scheduled shutdown with the most serious difficulty being the rupture of the 316 stainless steel slurry-preheater coil. It is suspected that stress corrosion cracking induced by the presence of halides was responsible for the rapid failure of the coil. Operational problems with the vacuum flash system, consisting primarly of plugged discharge lines, were the next most serious problem encountered.

A simplified flow diagram for the PDU was given in the Quarterly Progress Report No. 1 for April, May and June, 1974. A set of more complete flow diagrams is included in this report as Figures 1 through 9. Details of operation are discussed in the following sections corresponding to areas indicated on the flow diagrams.

1. Area 01A: Coal Crushing and Grinding

Operation of this area was satisfactory with no problems encountered.

2. Area 01B: Slurry Preparation and Pumping

Minor problems were encountered such as plugging in the slurry recirculating line, the drop chute and metering auger. The mixing propeller came off the shaft during one shakedown test. These difficulties were corrected.

Flow control of the circulating solvent stream is difficult because of variation in viscosity of the stream depending upon proportions of recycle and make-up solvent. This variability in viscosity also influenced operation of the slurry mix tank, and close adjustment of the flow rate was required to keep the contents of the tank uniform and pumpable. The orifice flow transmitter in the solvent supply line became plugged with fine solids accumulated over several weeks of operation.

3. Area 02: Slurry Preheating, Dissolving and Gas Separation

Use of hydrogen and carbon monoxide in the system remedied feed slugging problems between the dissolvers and the high pressure flash separator previously reported. The gas-liquid separators have been performing satisfactorily with only minor control problems.

The third shakedown run was terminated when a number of pinhole leaks were discovered in the high pressure tubing conducting vapors between separators S-1 and S-2. The tubing was replaced. Metallurgical examination of the failed section is still in progress.

The following run was terminated when the sandbath preheater coil became plugged with coke which formed when the slurry feed rate was reduced in an effort to keep the plant operating while the vacuum bottoms pump was being repaired in Area 04. When the preheater was disassembled and the sand removed, it was found that a number of the immersion heaters in use had failed because of hot spots which developed where the fluidized sand had caked on the heater sheaths. The coil and heaters were replaced.

Failure of heat tapes used for line tracing remained a problem. Irregularities of surface contour result in hot spots and early tape failure. Some loosening of fittings and flanges due to thermal cycling has become apparent. Several cone and thread fittings had to be tightened during downtime.

During shakedown test SDR-6, the tungsten carbide valve plug in a level control valve separated from its stem and remained seated. The plug blocked liquid outflow from S-3 which overflowed, filling S-4A, and resulted in the safety valve releasing hot solvent to the thermal-oxidizer collection system. Some of the solvent, however, backed up through vent lines into product storage drums. Piping modifications were made to prevent discharge into the product drums.

During Run SDR-7, the tungsten carbide valve seat in a control valve (LTV-3A) cracked and was wire drawn. A pressure surge caused the safety valve on S-1 to release. The safety valve trim was eroded by the coal slurry and preparations were being made for replacement when the E-1 preheater coil failed. The coil failure released 15 to 20 gallons of hot coal slurry plus 80 to 120 scf of hydrogen and carbon monoxide into the air-fluidized sand bath. The resulting fire was contained within the sand bath and an orderly shut down was conducted. Inspection of the coil after removal from the sand bath showed numerous circumferential cracks. Failure was apparently caused by stress corrosion cracking induced by chlorides. The coil was fabricated from type 316 stainless steel cold-drawn tubing. A replacement coil of similar 316 composition but made of annealed tubing is being procured.

4. Area 04: Liquid Separation and Solvent Recycle

Operational problems were encountered with the vacuum flash drum bottoms pump, the two vacuum pumps, the light ends column and thermal cycling of flange bolts. Run SDR-6 was terminated because a plug above the nuclear level gauge caused the vacuum flash drum to overfill and dump slurry into the recycle condensate receiver, S-7. Inspection of the disassembled unit revealed little deposition or erosion.

Light oils and water accumulating in the vacuum pump and in the vacuum pump oil caused considerable corrosion resulting in failure of pump bearings. Both vacuum pumps were rebuilt, and a cold trap employing a dry ice-trichloroethane bath was installed up-stream of the vacuum pump.

Some plugging of the light ends column occurred when a heavy tar-like material settled to the bottom of the reboiler and obstructed the outlet pipes. This material may have been entrained in vapors leaving the vacuum flash drum or may have been an integral part of the recovered solvent which settled out when the light ends were removed.

The Moyno pump, required for the removal of vacuum flash bottoms, was a constant source of trouble and was responsible for about 70 percent of the forced shut downs during the first shakedown tests. The drive shaft has sheared, leaks have developed in the seals, and the pump has lost suction; without fluid the rotor binds to the stator and "freezes". Several other mechanical problems developed, but were corrected. However, when reacted coal slurry was in the vacuum flash leg, operation of the pump was satisfactory until the coal slurry solidified because of low temperatures on the pump or the discharge line.

5. Area 05: Gas Purification and Recycle

Minor problems were encountered with the hydrogen and recycle gas compressors when both compressors automatically shut down for no apparent reasons. Chart recorders indicated stability of pressure which should have prevented the low pressure switch from shutting down the compressors. Operation has subsequently been normal.

Although the product gases are presently being discharged to the thermal oxidizer, the CO_2 scrubber was operated for a short period during one shakedown run. The flow transmitter on the sodium hydroxide solution was faulty and flow and level could not be controlled, with the unit eventually becoming overfilled. The transmitter was replaced.

B. PDU DESIGN AND CONSTRUCTION

1. Area 00: Buildings and Utilities

Cabinets, benches, lights and electrical services were installed in the plant control laboratory and the main disconnect switch in the Coal Preparation Building was changed to a location in the control room. All electrical work is now completed in this building.

Work on the extension of the mezzanine floor in the west bay of the Process Development Building was completed. This additional area will be used for the instrument panels for the Solid-Liquid Separation and the SRL Hydrogenation Units.

2. Area 02: Slurry Preheating, Dissolving and Gas Separation

Two stainless-steel level transmitters, S-4A and S-4B, were installed and piped into the system. Nine other stainless-steel level transmitters will be installed shortly. The stainless steel units will replace the presently installed carbon steel ones.

A safety override was designed and installed for the captive volume systems below separators S-1 and S-3. The override will prevent overlapping in the open position of the two slurry let-down valves in the event the control cam timer should slip.

3. Area 03: Solid Liquid Separation

Both benzene pumps were installed as were the instrument panels. Electrical work is being delayed pending receipt of junction boxes, stop-start stations, and motor starters. The P&I flow diagram was reviewed and the control system revised to include additional control valves.

4. Area 04: Liquid Separation and Solvent Recycle

The Elf turbine meter and totalizer were installed. Major modifications were made to the vacuum flash system. A gear pump was installed at the base of the vacuum flash leg to remove light vacuum bottoms until sufficient product accumulates to provide sufficient head and/or viscosity for operation of the Moyno pump. The Moyno pump was rebuilt with a heavier drive end and connecting rod installed. A reversing switch was added.

Discharge lines were modified and shortened. Improved heat tracing and temperature controls were installed.

A small knockout pot was installed in the gas line from Separator S-4A to the Light Ends Column to protect the back pressure rotameters from oil entrained with the gas leaving Separator S-4A.

5. Area 06: SRL Hydrogenation

The SRL charge pump was installed. The shell for the sand batch preheater was completed and installed. Electrical band heaters are being attached to the shell.

6. General

A trial yield period of 12 hours was made during a shakedown run to allow checking of mass balance procedures and methods of data collection. A computer program is being developed to aid in reduction and presentation of PDU data. Additional sample points have been added to provide needed mass balance data.

An Antek gas chromatograph and recorder was received and is being prepared for use.

Safety relief valves have presented a reseating problem. One was returned to the manufacture for inspection and suggestions on how to avoid the problem. The seat material was replaced with a harder stainless steel which may solve the reseating problem.

Instruments are being installed in the west bay instrument panel. The additional transformers for control of the heating tapes have been installed.

C. EXPERIMENTATION

Fourteen batch autoclave experiments were performed in the investigation of process conditions for liquefaction of lignite. Four of these used lignite pretreated with phenol as feedstock under otherwise standard liquefaction test conditions of 30 minutes at 752°F reaction temperature, a 2:1 weight ratio of solvent to MAF lignite, and a 1:1 molar ratio of hydrogen to carbon monoxide at 1000 psig initial pressure. Product removal was also standard with the autoclave being cooled to 400°F, and the gaseous and the solid-liquid products discharged at that temperature. Gaseous products were released slowly into a series of condensate cold traps ranging from room temperature to a -70°C. The noncondensibles were collected in a 200-liter rubber gas bag and analyzed for hydrogen sulfide, specific gravity, and the volume determined by discharge of the gas bag contents through a wet test meter. The other gaseous components were determined using gas-solid chromatography. Liquid-solid products were discharged from the autoclave under a 100 psi carbon monoxide pressure, and the ejected reactor slurry was suction-filtered at 400°F giving a filtrate containing SRL plus solvent and a cake containing the lignite ash with the unreacted lignite.

The four SRL hydrogenation tests reacted approximately 284 grams each of solvent refined lignite from Spencer Chemical Company with hydrogen at 1800 psi initial pressure in the presence of 217 grams of freshly sulfided Ni-Mo catalyst. The reactants were heated to 450°C (842°F) at a rate of 5°F/min, and held at that temperature for two hours. The autoclave was cooled to 400°F, and the gases removed using standard procedures. After removal of the gases, the reactor was cooled overnight to room temperature and the solid-liquid products discharged as before. As these tests were performed in conjunction with the hydrogenation work in the chemistry department, results are discussed under the section on University Supporting Research.

The four tests concerned with reactivity of lignite stored under various conditions completed the series of samples stored for 70 weeks under various conditions to simulate possible storage modes. Standard operating procedures were followed.

In the two tests for direct hydrogenation of lignite, no solvent was added and synthesis gas (1:1 molar ratio of hydrogen to carbon monoxide), at 1000 psig initial total pressure was used as the hydrogenation agent. Water was added to the pulverized lignite to attain the standard moisture content before charging. Otherwise standard operating procedures were used.

Detailed equipment descriptions and operating and analytical procedures for operation of the batch autoclave were given in the Quarterly Technical Progress Report No. 1 for April, May and June 1974. Usually, Gulf FS120, a carbon black feedstock, is used as a solvent. Pulverized lignite is processed in batches of about 200 grams on a MAF basis with the moisture content adjusted to 33 percent on an ash-free basis by addition of small quantities of water if required. Material balances and analytical data for the direct hydrogenation and reactivity test series are presented in Table 1; Parts 1 and 2, Appendix A, Section V. Discussion of results follow:

1. Direct Liquefaction Without Solvent

The conversion and yields on a 100 percent recovery basis are presented below for normal solvent liquefaction with syngas (using the average of 19 replicate runs made in the solvent recycle series), direct liquefaction using CO (Run 591), and direct liquefaction using syngas (Runs 597 and 598):

Run No.	Avg. of 19	591	597	598
	Solvent Liquefaction Runs with Syngas	Direct Liquefaction (no solvent)		
		with CO	with Syngas	
Yields, Wt% MAF Coal				
Gas	34.0	49.1	28.6	37.2
Net Liquid	64.6	63.0	34.3	53.5
Unconverted	11.4	6.1	39.3	16.7
H_2O & Ash	-10.0	-18.1	-2.3	-8.1

The data from the replicate direct liquefaction tests using syngas (Runs 597 and 598) do not agree as well as those for the usual solvent liquefaction tests. However, the liquid yields are significantly lower and the unconverted lignite significantly higher than in either of the other tests in comparison. It appears that direct liquefaction using pure CO as the reducing gas is much more attractive than using syngas (CO and H_2). Net SRL yields were similar for the test using CO without solvent and syngas with solvent.

2. Reactivity of Lignite After Storage

Pertinent data for the Reactivity Series tests after 70 weeks of storage are presented below along with baseline data obtained at the start of the series. No further tests are planned for this series; a comparison of all tests will be presented in a future report.

Run No.	504 & 505 (average)	592	593	594	595
Description	Baseline	N_2 Storage		Air Storage	H_2O Storage
Yields, Wt% MAF Coal					
Gas	36.8	47.1	44.1	49.2	47.2
Liquid	66.4	63.6	61.5	58.9	57.3
light oil	(24.4)	(23.5)	(18.3)	(13.8)	(18.4)
net SRL	(42.0)	(40.1)	(43.2)	(45.1)	(38.9)
Unconverted	8.8	6.2	7.4	7.3	9.3
H_2O	-12.0	-16.9	-13.0	-15.4	-13.8
Filterability Index*	12.0	10.1	15.0	12.7	13.7

*Filterability index is the percentage of total output filtrate retained in the filter cake, residue, and pot residue.

While the unconverted material was essentially the same independent of storage method for 70 weeks compared to the baseline tests, yield of gaseous products seemed to increase while yields of SRL did not change appreciably except when storage was under water.

3. Phenol-Pretreatment Tests

The initial study of phenol pretreatment on liquefaction characteristics of lignite was completed. Treatment of lignite with a phenol-water slurry prior to liquefaction had been suggested as a method for conditioning of lignite prior to liquefaction to improve solubilization characteristics. Full moisture lignite was treated with aqueous phenol, ranging from 30 to 100 percent phenol. In one instance, phenol was added directly to the liquefaction autoclave just prior to liquefaction. Generally, following the phenol treatment, the lignite was washed with warm water and followed by a light rinse with ambient temperature water to remove excess phenol. After the rinse, the lignite was processed using normal liquefaction procedures. Experimental results and analytical data for the complete series are given in Appendix A, Table 1, Parts 1 and 2, runs PTR-1 through PTR-7. Test PTR-1 was a standard liquefaction test to serve as baseline data for this series. Experimental conditions and yield data on a MAF lignite basis are summarized in the following:

Run No.	PTR-1	PTR-2	PTR-3	PTR-4	PTR-5	PTR-6	PTR-7
Yields, Wt% MAF Lignite							
Net Gas	47.9	39.4	40.4	42.1	56.3	37.0	41.4
Net Liquid	64.2	52.0	62.2	56.2	62.9	68.6	77.4
Unconverted	6.1	10.1	8.8	6.2	5.5	7.1	6.7
Net H_2O+Ash	-18.2	-1.5	-11.4	-4.5	-24.7	-12.7	-25.5

Conditions: PTR-1, Base; PTR-2, 100% Phenol treat, 212°F wash, light rinse; PTR-3, 100% phenol treat, 140°F wash, light rinse; PTR-4, 100% phenol treat, Ambient wash, light rinse; PTR-5, phenol treat, crystals added; PTR-6, 100% phenol treat, light rinse; PTR-7, 30% phenol treat, light rinse.

The data indicate little or no improvement in basic yields as a result of phenol treatment. Total conversion as indicated by the extent of unconverted lignite is similar to the differences obtained in replicate runs. The test with phenol pretreatment using a 30 percent slurry of phenol followed by a light rinse, did indicate a higher yield of net liquids and higher loss of net water plus ash. However, values may be faulty because of analytical difficulties.

A gas-liquid chromatographic method was developed to determine the concentration of phenol in both aqueous and non-aqueous media. A Beckman GC-2 gas chromatograph instrument was used, employing a ¼-inch o.d. x 4-foot long stainless steel column. The column was packed with Chromosorb W solid support loaded with 5.5 wt percent of Carbowax 6000. The column and detector were controlled isothermally at 155°C; the injection port was maintained at 229°C. Bridge current used was 350 ma. Using helium as the carrier gas with a flow rate of 27 cc/min, phenol in the unknown sample was resolved in approximately 6.5 minutes. This method is currently in use for determining phenol in water and light oil samples.

4. Variability of Lignite Analyses

Over a period of two years, analytical data have been accumulated on the lignite charged to the batch autoclaves in the laboratory liquefaction program. Lignite source was North American Coal Company's Indian Head Mine near Zap, North Dakota. Bulk samples were obtained and carefully stored with small portions removed, pulverized and used in liquefaction tests. Partial proximate and ultimate analyses were made for each run. Data from 67 analyses were used to calculate the variability introduced by different samples and different chemists. Results are summarized in the following:

Proximate Analysis, Wt%	Average	Standard Deviation
Moisture	28.5	1.82
Volatile Matter	31.4	1.70
Ash	6.67	0.683
Ultimate Analysis, Wt% MAF		
Carbon	64.5	1.45
Hydrogen	4.27	0.261
Sulfur	0.873	0.234
Nitrogen*	0.798	0.129

* Only 12 values used in calculation

The results indicate a relatively limited variability in the analyses. The standard deviations reported are for a single value, not of the average. Moisture content is low as there was loss during pulverization to -100 mesh and preparation of sample. Both moisture and ash contents are generally considered to be the more variable of the analyses. More uniformity is exhibited in the moisture-ash-free ultimate analyses reported. Of these, sulfur is usually the most variable.

5. University Supporting Research - Conversion of Solvent Refined Lignite to Liquid Fuels.

During the present reporting period, nine hydrogenation experiments were completed in investigation of hydrogenation of solvent refined lignite (SRL). Detailed experimental procedures were described in the Quarterly Technical Progress Report No. 3 for October, November and December, 1974. Unless specified otherwise, the standard reaction conditions were two hours residence time at a reaction temperature of 450°C with an initial hydrogen pressure of 1800 psig; approximately 75 grams of SRL or SRC were charged.

Runs 72 and 75 investigated tungsten disulfide as the catalyst. In Run 72, 7.5 g of WS_2 was added together with the usual quantity of $Ni\text{-}Mo\text{-}Al_2O_3$ catalyst while in Run 75, 37.5 g of WS_2 alone was used.

Runs 73 and 74 were performed in the Project Lignite Laboratory where a one-gallon high pressure autoclave was available. General procedures for these two runs were similar to those previously described for Run 70 except that the glass helices were removed from the cold traps for the light boiling liquids. This modification improved the recovery of the gasoline boiling range fraction from 20.1 g in Run 70 to 40.2 g.

For Runs 76 and 78, the solvent was 150 ml of tetralin for 75 g of SRL. A reaction temperature of 450°C was employed and the reaction time was two hours. An initial hydrogen pressure of 2500 psi was used and 37.5 g of presulfided $Ni\text{-}Mo\text{-}Al_2O_3$ added.

Runs 77 and 79 were without solvent. For Run 77, the catalyst was conditioned in a hydrogen atmosphere for four days while for Run 79, the catalyst was not hydrogen conditioned nor exposed to air after presulfiding. For Run 80 test conditions were similar to those of Run 79 with the exception that SRC was used as feedstock rather than SRL. More detailed experimental conditions and experimental results are presented in Tables 3 and 4, Part VI, Appendix B. Table 5 in Appendix B, summarizes the NMR analyses of the distillation fractions.

The following observations are made concerning the hydrogenation experiments:

a. The addition of WS_2 with the usual quantity of $Ni\text{-}Mo\text{-}Al_2O_3$ catalyst did not influence conversion compared to tests with $Ni\text{-}Mo\text{-}Al_2O_3$ alone.

b. WS_2 by itself is not as good a catalyst as the presulfided $Ni\text{-}Mo\text{-}Al_2O_3$. Unusually high concentrations of aromatics were found in the products.

c. Larger scale batch reactor tests indicate similar conversions with the exceptions that yields of gaseous products were increased (about 8 percent) and the gasoline yield reduced (about 10 percent).

d. Hydrogen conditioning of catalyst did not influence overall conversion.

e. The hydrogenation test with SRC indicated that conversions are essentially similar to those of SRL under the same test conditions. Over 20 percent more ammonia was produced during liquefaction of SRC because of a higher initial nitrogen content.

f. Using tetralin as solvent, mixing during the hydrogenation reaction is important even in the presence of large quantities of solvent. For a nonstirred test, conversion was 80 percent, and for a stirred test, 88 percent.

D. ECONOMIC STUDIES

Since the inception of the Project Lignite program, economic studies of lignite conversion have been conducted in cooperation with members of the UND Economics Department. Two of the studies that were recently concluded are reported herewith.

1. A Comparative Economic Evaluation Of Two Plants For The Solvent Refining Of Lignite

This is a comparative economic evaluation of two kinds of plants for processing North Dakota lignite into a composite product consisting mainly of a clean burning boiler fuel called Solvent Refined Lignite (SRL) and a light liquid resembling crude oil. The first plant is a straightforward modification of the Solvent Refined Coal process (SRC) described in OCR R&D Report No.

53, Interim Report No. 1 (1970), hereinafter referred to as the Standard Solvent Refined Lignite Plant (SSRL). The second plant is also a modified version of the SRC process but with certain important alterations, the most significant of which is the substitution of a Lurgi gasification plant for the natural gas reforming plant to produce the reducing gas for liquefaction. This latter facility will be referred to henceforth as the Solvent Refined Lignite with Lurgi Plant (SRLL).

The SSRL has a processing capacity of 35M tons of lignite per stream day from which 8,850 tons of SRL and 2,830 tons of light oil are produced. The SRLL has a processing capacity of 39M tons per day from which 8,850 tons of SRL, 2,976 tons of light oil, and 3,432 tons of synthesis gas are produced.

The first plant has three major operating costs (1) lignite, (2) synthesis gas, and (3) labor; while lignite and labor are the major cost items in the second process. In each case lignite is purchased for \$2.25/ton and is assumed to contain 35% moisture and 6% ash. Labor is hired at \$5.00/hr and make-up gas is purchased for 60 ¢/MM Btu.

Total investment for each plant is estimated at \$237 million. Although capital costs are similar, annual operating costs show a substantial divergence with total cost being estimated at \$81 million for SSRL and \$57 million for SRLL before by-product credits. Electricity and sulfur are the chief by-products whose combined value ranges from \$3.5 to \$2.4 million for the two processes respectively.

The investments are assumed to be 100% equity financed and to require an internal rate of return of 12%. Discounting the foregoing data at this rate yields a product selling price of \$28.51/ton for the SSRL and \$22.45/ton for SRLL.

a. Process Description

Lignite is crushed, ground, and combined with a recycle solvent to form a slurry. The slurry mixed with recycle gas and some make-up synthesis gas (CO and H_2) is heated and fed into a high pressure dissolver where it reacts for about thirty minutes. The products from the reactor go into a separator for the splitting off of gases and light liquids. The gas is then purified and subsequently divided into two streams one of which is recycled to the dissolver. The slurry itself is filtered to remove unconverted coal and other mineral matter with the resulting filtrate being distilled into SRL, light oil, and the recycle solvent. Thus the processes are basically alike except in their handling of synthesis gas. In the SRLL facility this make-up gas is supplied by a Lurgi gasification component whereas in SSRL this gas requirement is met by reforming purchased natural gas.

b. Investment

The total investments for the SSRL and SRLL plants are shown in Table 6 and Table 7. The investment figures encompass total plant costs, capitalized start-up costs, return on equity during construction, and working capital but exclude plant site costs, water transport investment, and contingencies. Investment requirements for the basic SRL process were computed by modifying and scaling up the SRC process steps set forth in the previously mentioned OCR report. Additional modifications were made in estimating the costs of SRLL by determining Lurgi gasification requirements from BOM Report No. 7408 (1970) and by adapting and scaling down the Lurgi system presented in BOM Report 70-21, Case III (1970).

Plant construction, including site preparation and support buildings, is assumed to occur over a three-year period in three equal installments. In addition it is assumed that start-up costs equal to 7% of total plant cost are incurred during the final year of construction and are expensed.

c. Operating Costs

Operating costs are based on prices which might have prevailed in western North Dakota in mid-1974. Input requirements were obtained by adapting and scaling the relevant figures presented in the above-mentioned reports.

Lignite requirements for the two processes have already been cited. In the case of SSRL another major input not supplied internally is the natural gas which must be purchased at the previously mentioned rate. The amount required is estimated at 5.5 billion Btu per hour.

The operating labor requirement is 200 men for SSRL (Table 6) and 221 men for SRLL (Table 7). Maintenance costs were derived as percentages of the individual process investments and allocated to materials and labor by a pro-rata formula used in OCR Report No. 53. Other direct production costs include water, chemicals and catalysts. Plant overhead and general expenses include services indirectly related to operation of the plant, namely payroll taxes, fringe benefits, and general administrative expenses.

Depreciation is included in fixed charges along with property tax and insurance. Depreciation is based on the straight line method with an assumed plant life of 20 years. Total operating cost before by-products credits is $81 million for SSRL (Table 6) and $57 million for SRLL (Table 7).

By-product credits are taken for electric power and sulfur, Excess electricity is valued at 0.5¢/KWH and a credit for sulfur is taken at $15/LT. Total by-product credits are $3.5 and $2.4 million for SSRL and SRLL respectively, yielding net operating costs of $77 and $55 million annually.

d. Product Selling Price

A required selling price for the composite good is arrived at by discounting the cost data presented in Table 6 and Table 7 on the basis of a required rate of return of 12% using the Discounted Cash Flow (DCF) method. This base price is then prorated between SRL and light oil on a 4:11 basis, where a typical ton is assumed to contain 3 times as much SRL as light oil.

The price of the composite good produced by the SSRL process is $28.51/ton (Table 8) or 82.9¢/MM Btu. The 4:11 formula yields a selling price of $19.83/ton for SRL and $54.54/ton for light oil.

Under the SRLL process the product price is $22.45/ton or 65.1¢/MM Btu. The price of SRL is $15.62/ton and $42.95/ton is the price of light oil.

The results of an analysis of the sensitivity of product price to changes in basic parameters (DCF, coal cost, and labor cost) are presented in Table 8. As one would expect, the data indicate that product price is relatively more sensitive to changes in the discount rate and the cost of coal than to changes in labor cost.

e. Conclusions

Significant savings are realized by producing the required make-up gas from lignite rather than from natural gas purchased from a commercial supplier. This at least is true where the SRC process utilizes lignite as a feedstock. Whether or not this holds in general cannot be said at this time without further investigation. In the process in question, however, the savings are generated by substituting a low-cost source of synthesis gas for supplies usually produced from gas purchased at 60¢/MM Btu. This reduction in operating cost translates directly into a lower selling price for the final product.

2. An Economic Evaluation Of A Lurgi Gasification System For Producing Low-BTU Gas From Lignite.

This is an economic evaluation of a plant designed to produce 121 million standard cubic feet per day of 350-Btu gas. Plant requirements were estimated from operating data for the slagging, fixed-bed gasifier published in BOM Report No. 7408(1970) and by adapting the Lurgi system presented in BOM Report 70-21, Case III (1970). Major components of the system envisioned here are coal preparation, gasification, purification, and sulfur recovery. The plant has a processing capacity of over 3M tons of lignite per stream day from which the low-Btu gas is produced.

Total investment is estimated at $21 million and assumed to be 100% equity financed. A rate of return of 12% is required. Coal and labor are the major operating costs with lignite being purchased for $2.25/ton and labor hired at $5.00/hr. Annual operating costs are $5.6 million before by-products whose combined value is $65 thousand per year. When the foregoing data are discounted at the 12% rate a selling price of 68¢/MM Btu is obtained for the product.

a. Lignite is crushed and fed into a slagging gasifier where it reacts with steam and oxygen at a pressure of 400 PSIG. Low-Btu gas is the major product leaving the gasifier but a substantial amount of tar is also produced. The products of the gasification step are separated before the gas is purified and sulfur is removed.

The gasifier is similar to the one developed by Lurgi over thirty years ago. In the Lurgi process coal, steam, and oxygen react at high pressure in order to maximize the methane content of the producer gas. This feature is retained in the slagging unit developed by the Bureau of Mines. The chief difference lies in the handling of the ash. While ash is removed by a mechanical grate in the original Lurgi it forms a molten slag in the version used here. However, the higher temperature results in somewhat less methane formation.

b. Investment

The total investment of $24,984,000 is shown in more detail in Table 9. The investment figures encompass total plant costs, capitalized start-up costs, return on equity during construction, and working capital but exclude plant site costs and contingencies.

Plant construction, including site preparation and support buildings, is assumed to occur in equal installments over a three-year period. In addition it is assumed that start-up costs equal to 7% of total plant cost are incurred during the final year of construction and are expensed.

c. Operating Costs

Operating costs are based on prices which might have prevailed in western North Dakota in mid-1974. Input requirements were obtained by adapting and scaling the relevant figures presented in the above-mentioned reports.

The operating labor requirement is 20 men involving an outlay of $200 thousand a year. Maintenance costs were derived as percentages of the individual process investments and allocated to materials and labor. Other direct production costs include water, chemicals, and catalysts. Plant overhead and general expenses include services indirectly related to operation of the plant; namely payroll taxes, fringe benefits, and general administrative expenses.

Depreciation, property tax and insurance are included in fixed charges. Depreciation is based on the straight line method with an assumed plant life of 20 years. Annual operating cost before by-product credits is $5.6 million including $1.0 million depreciation.

By-product credits are taken for steam and sulfur. Excess steam is valued at 50¢/MM Btu and a credit for sulfur is taken at $15/LT. Total by-product credits are $65 thousand yielding net operating costs of $5.5 million annually.

d. Product Selling Price

A required selling price for the joint-product is arrived at by discounting the cost data presented in Table 9 on the basis of a rate of return of 12% using the Discounted Cash Flow (DCF) method. This procedure yields a selling price of 68¢/MM Btu for the gas and tar.

The results of an analysis of the sensitivity of product price to changes in basic parameters (DCF rate, coal cost, and labor cost) are presented in Table 10. As one would expect, the data indicate that product price is relatively sensitive to changes in these values.

V. CONCLUSIONS

A. PDU OPERATIONS

1. Seven shakedown runs were started in the liquefaction pilot plant and all were terminated prior to scheduled shutdown.

2. Operational and minor equipment problems were correctible. However, a serious difficulty was the sudden rupture at operating temperature and pressure of the stainless steel slurry preheater coil.

3. Coil failure was caused by stress corrosion accelerated by the presence of halides.

B. PDU DESIGN AND OPERATION

1. Since most of the PDU is in operational status, major emphasis in construction activities was given to the solids liquid separation and SRL hydrogenation areas.

2. Pumps and instrumentation panels were installed in the solids liquid separation area although back ordering of electrical supplies has delayed completion.

3. Substantial progress was made in the SRL hydrogenation area where the charge pump and the shell for the preheater were installed.

4. A trial yield period during a shakedown run with the PDU for obtaining material balances established data collection procedures and indicated some necessary modification.

C. EXPERIMENTAL

Fourteen batch autoclave tests were completed in the investigation of process conditions for liquefaction of lignite.

1. In direct liquefaction tests without solvent, yields of liquid products were higher and those of unconverted lignite lower when using carbon monoxide than when using synthesis gas.

2. Net SRL yields were similar for tests using carbon monoxide without solvent and synthesis gas ($CO + H_2$) with solvent.

3. Storage of lignites for 70 weeks in air, in a nitrogen atmosphere or under water did not appreciably effect overall conversion although yields of gaseous products seemed to be higher than for base line tests.

4. Pretreatment of lignite with phenol prior to normal liquefaction did not improve basic yields.

In University Supporting Research on catalytic hydrogenation of SRL, nine tests were completed.

1. Addition of WS_2 to $Ni-Mo-Al_2O_3$ did not effect conversions.

2. Hydrogen conditioning of catalyst did not influence overall conversions.

3. Hydrogenation of SRC and SRL under similar conditions gave essentially the same conversions.

4. Tests on a larger scale indicated that conversions were not appreciably influenced except for possible increase in gaseous components.

5. Mixing during hydrogenation is necessary for high yields even with large quantities of solvent.

D. ECONOMICS

Substitution of a slagging Lurgi gasifier for a natural gas reformer to produce reducing gases for lignite liquefaction results in a decrease in the required price for solvent refined lignite from 83¢/MM Btu to 65¢/MM Btu.

A small self-sufficient plant to produce low Btu (350 Btu/CF) reducing gas from lignite using a slagging Lurgi gasifier will require a product selling price of 68¢/MM Btu.

VI. APPENDICES

A. LABORATORY EXPERIMENTATION

Table 1 - Solution Hydrogenation of Phenol Treated Lignite

Part 1: Material Balance and Yield Data
Part 2: Analytical Data

Table 2 - 70-Week Reactivity Tests and Non-Solvent Hydrogenation of Lignite.

Part 1: Material Balance and Yield Data
Part 2: Analytical Data

B. UNIVERSITY SUPPORTING RESEARCH

Table 3 - Hydrogenation of SRL in One-Liter and One-Gallon Autoclaves

Table 4 - Hydrogenation of Solvent Refined Lignite, Distillation Fractions.

Table 5 - NMR Analyses, Hydrogenation Experiments, Ar-H/Al-H Ratios of Distillation Fractions.

C. ECONOMICS

Table 6: Economic Summary - Standard SRL Process
Table 7: Economic Summary - SRL with Lurgi
Table 8: Product Selling Price - SSRL and SRLL Processes
Table 9: Economic Summary - Slagging Lurgi Gasification
Table 10: Product Selling Price - Slagging Lurgi Gasification

D. PDU FLOW SHEETS

Figure 1: Coal Crushing and Grinding, Area 01A.
Figure 2: Slurry Preparation and Pumping, Area 01B.
Figure 3: Slurry Preheating and Dissolving, Area 02A.
Figure 4: Gas Separation, Area 02B.
Figure 5: Solid-Liquid Separation, Area 03.
Figure 6: Solvent Recycle, Area 04A.
Figure 7: Liquid Separation, Area 04B.
Figure 8: Gas Purification and Recycling, Area 05.
Figure 9: SRL Hydrogenation, Area 06.

TABLE 1 - PART 1: MATERIAL BALANCE AND YIELD DATA

SOLUTION-HYDROGENATION OF PHENOL TREATED LIGNITE

Run No.	PTR-1 (4)	PTR-2	PTR-3	PTR-4	PTR-5 (6)	PTR-6	PTR-7
	Baseline Test	100% Phenol Treat.			See Footnote	100% Phenol Treat	30% Phenol Treat
		212°F H_2O wash	140°F H_2O wash	H_2O Pretreat			
		Lt. Rinse				Lt. Rinse	
Test Conditions							
Time, hrs	0.5	0.5	0.5	0.5	0.5	0.5	0.5
Avg. Temp., °F	752	752	752	752	752	752	752
Max. Press, psia	3435	3465	3435	3145	3385	3065	3335
Gas Charged	CO/H_2						
Solvent	UNDFS120-75						
Lignite	N.A. 0.132 In.						
Material In, grams							
Coal	323.9	297.6	287.4	227.6	233.7	214.2	205.5
H_2O	0.0	6.9	13.1	6.5	20.2	27.9	26.3
Solvent	399.6	400.4	400.2	400.3	399.5	399.8	400.0
Gas	150.9	155.0	148.0	152.8	151.2	153.6	156.6
Phenol	0.0	11.6	15.6	81.2	60.6	70.3	30.4
Total	874.4	871.5	864.3	868.4	865.2	865.8	818.8
Material Out, grams							
Filtrate	448.7	431.8	453.2	484.9	473.2	469.7	451.0
Filter Cake	61.4	67.9	58.4	42.5	37.4	51.1	33.2
Residues	39.9	35.1	38.1	17.5	20.1	23.0	19.1
Gas	242.6	229.2	221.3	217.5	223.5	207.6	210.9
H_2O	63.2	88.9	61.6	37.4	68.9	35.3	40.1
Light Oil	2.4	4.4	6.0	37.1	20.9	40.6	41.9
Total	858.2	857.3	838.6	836.9	844.0	827.3	796.2
% Recovery	98.2	98.4	97.0	96.4	97.5	95.6	97.2
Yields 100% Recovery Basis							
Products, grams							
Light Oil (1)	20.8	27.0	38.0	102.9	69.0	114.4	81.2
Solvent	395.2	380.6	384.0	393.7	402.3	398.6	396.3
Vac. Btms	112.9	107.3	117.1	82.1	75.8	75.2	65.6
Total Liquid	528.9	514.9	539.1	578.7	547.1	588.2	543.1
Net Liquids (2)	129.3	102.9	123.3	97.2	87.0	118.1	112.7
Net Gas	91.7	77.9	80.1	72.9	77.9	63.7	60.3
Net H_2O (3)	-38.5	-4.2	-27.8	-9.7	-33.7	-22.4	-39.3
Net Ash	1.9	1.3	5.2	2.0	-0.4	0.5	2.1
Unconverted Coal (MAF)	12.2	20.0	17.5	10.6	7.7	12.3	10.0
Yields Wt% MAF Coal							
Net Gas	47.9	39.4	40.4	42.1	56.3	37.0	41.4
Net Liquid	64.2	52.0	62.2	56.2	62.9	68.6	77.4
Unconverted	6.1	10.1	8.8	6.2	5.5	7.1	6.7
Net H_2O + Ash	-18.2	-1.5	-11.4	-4.5	-24.7	-12.7	-25.5
Filtrate Composition (Wt%)							
Light Oil	3.4	3.8	5.5	11.6	8.3	12.8	7.1
Solvent	75.1	75.1	72.4	73.2	77.2	73.4	79.7
Vacuum Btms	21.5	21.1	22.1	15.2	14.5	13.8	13.2
Net Liquid Composition (Wt%)							
Net Light Oil	10.3	13.7	19.2	59.5	49.9	66.4	55.3
Net SRL	53.9	38.3	43.0	-3.3	13.0	2.2	21.6
Solvent Recovery							
Wt%	97.1	93.5	93.1	94.8	98.2	95.3	96.3
Wt% (100% Rec. Basis)	98.9	95.1	96.0	98.4	100.7	99.7	99.1

* Footnotes on following page.

FOOTNOTES ON PTR SERIES TESTS:

(1) Light oil yield is corrected for initial grams of phenol in input coal and grams of phenol analyzed in water phase (product water).

(2) Net liquids yield is total liquids [corrected light oil yield + solvent + SRL - (grams of input solvent + grams of phenol in coal)].

(3) Net H_2O yield is grams input H_2O - grams output H_2O, corrected for grams phenol analyzed in water phase.

(4) PTR-1 is a baseline test for this test series; the coal was not pretreated with phenol nor was phenol added prior to liquefaction.

(5) H_2O pretreatment means that the coal was prewashed with RT water prior to treatment with phenol.

(6) PTR-5 was conducted to investigate the effects of adding phenol crystals to the input coal at the time of charging to the autoclave. This is in contrast to Runs PTR-2, 3, 4, 6 and 7 which the lignite was pretreated with aqueous phenol, ranging from 30 Wt% to 100 Wt% phenol, at elevated temperature; following treatment, the coal was rinsed lightly with water to remove a portion of the phenol.

TABLE 1 - PART 2: ANALYTICAL DATA

SOLUTION-HYDROGENATION OF PHENOL TREATED LIGNITE

	Phenol - Treatment Series						
Run No.	PTR-1	PTR-2	PTR-3	PTR-4	PTR-5	PTR-6	PTR-7
Test Conditions							
Time, hrs.	0.5	0.5	0.5	0.5	0.5	0.5	0.5
Avg. Temp., °F	752	752	752	752	752	752	752
Max. Press., psia	3435	3465	3435	3145	3385	3065	3335
Gas Charged	CO/H_2						
Solvent	UNDFS120-75						
Lignite	North American Lignite - 0.132 Inch Particles						
Analytical Data							
1. Gas Analysis, Mol%							
H_2	42.6	38.0	41.6	40.4	43.0	40.6	41.1
CH_4	3.3	2.5	2.6	2.5	2.6	2.2	2.3
CO	20.8	31.0	30.6	31.5	23.8	33.1	34.3
C_2H_6	1.0	0.9	0.9	0.7	0.7	0.7	0.6
CO_2	31.7	27.0	23.8	24.3	29.4	23.0	21.2
H_2S	0.6	0.6	0.5	0.6	0.5	0.4	0.5
Gas Specific Gravity @ RT	0.7968	0.7642	0.7464	0.7472	0.7409	0.7268	0.7049
2. Input Coal							
Volatile Matter, Wt% (3)	29.11	33.99	33.58	40.41	29.82	46.52	37.87
Ash, Wt%	6.26	5.00	4.51	4.75	6.68	4.63	3.58
Moisture, Wt% (3)	31.60	30.11	28.69	30.28	34.13	25.34	31.23
Carbon, Wt% (1)	66.46	63.61	65.54	66.48	64.73	70.64	69.15
Hydrogen, Wt% (1)	4.09	4.56	4.56	4.52	4.31	4.38	4.62
Sulfur, Wt% (1)	0.66	1.19	0.90	2.28	1.20	0.95	0.48
Nitrogen, Wt% (1)	0.96	0.90	0.87	0.80	0.90	0.74	0.72
Phenol, Wt% of Total	---	3.75	5.14	26.30	20.60 (4)	24.70	12.90
3. Input Solvent							
Ash, Wt%	0.01	---	---	---	---	---	---
Carbon, Wt%	88.63	88.55	87.51	90.21	89.33	88.56	89.06
Hydrogen, Wt%	7.98	8.08	7.98	8.04	7.98	7.95	8.09
Sulfur, Wt%	2.28	2.44	2.35	2.34	2.31	2.35	2.27
Specific Grav. 60/60	1.058	1.050	1.054	1.054	1.054	1.056	1.055
Brookfield Visc, cp (2)	---	61.0	67.0	71.0	76.5	76.5	75.0
IR Ratio	0.20	0.20	0.21	0.20	0.21	0.21	0.20
4. Coal-Solvent Slurry							
5. Cake							
Ash, Wt%	26.54	20.52	23.39	30.21	33.52	22.54	26.85
Carbon, Wt%	63.82	68.11	65.92	61.64	58.32	67.36	62.49
Hydrogen, Wt%	4.86	4.85	4.81	4.38	4.36	5.27	4.71
Sulfur, Wt%	2.82	2.45	2.99	3.35	3.45	2.50	2.94
Pyridine Sol., Wt%	76.38	71.06	72.73	71.33	73.31	76.34	67.83

(1) Calculated on a moisture-phenol-free basis

(2) Brookfield Viscosity measured at approximately 23°C.

(3) Moisture is determined by oven-drying the sample in air for one hour at 105°C; with phenol present in the coal, the moisture value includes phenol loss. Likewise, the volatile matter value includes the remainder of the phenol.

(4) Phenol was added to coal after the coal tests were conducted.

TABLE 1 - PART 2 CONT.: ANALYTICAL DATA

SOLUTION-HYDROGENATION OF PHENOL TREATED LIGNITE

Run No.	PTR-1	PTR-2	PTR-3	PTR-4	PTR-5	PTR-6	PTR-7
6. Filtrate							
Ash, Wt%	0.03	0.02	0.01	0.02	0.0	0.0	0.0
Carbon, Wt%	87.40	88.66	86.84	86.84	88.67	86.91	87.89
Hydrogen, Wt%	7.63	8.02	7.44	7.64	7.60	7.70	7.82
Sulfur, Wt%	1.75	1.87	1.79	1.82	1.79	1.83	1.97
Specific Gravity, 60/60	1.096	1.101	1.084	1.075	1.072	1.073	1.070
Brookfield Visc, cp (2)	2058	1357	1091	330	399	425	410
Blackness, abs/g @ 550 mu	5.29	5.51	5.03	2.72	2.68	3.36	2.98
7. Vacuum Btms							
Ash, Wt%	0.32	0.16	0.12	0.42	0.37	0.59	0.28
Carbon, Wt%	86.96	86.87	86.59	86.36	86.84	86.81	86.96
Hydrogen, Wt%	6.13	5.79	5.74	5.90	6.15	5.80	6.03
Sulfur, Wt%	0.54	0.63	0.50	0.51	0.50	0.54	0.50
Melting Pt., °F	290	298	324	309	301	305	295
8. Residue							
Ash, Wt%	13.38	6.65	12.07	18.04	10.79	6.37	6.74
Carbon, Wt%	78.66	81.91	76.97	75.60	80.16	82.04	82.17
Hydrogen, Wt%	6.80	6.75	6.39	6.27	7.07	6.88	6.95
Sulfur, Wt%	2.07	1.99	1.92	2.04	2.36	1.93	2.07
Pyridine Sol, Wt%	96.16	87.57	85.81	87.93	95.37	88.83	90.03
9. Light Oil							
Carbon, Wt%	80.87	84.20	76.33	64.67	66.72	62.62	65.52
Hydrogen, Wt%	9.97	11.77	8.09	7.70	8.41	7.33	7.55
Sulfur, Wt%	0.52	0.24	0.32	0.13	0.21	0.09	0.13
Phenol, Wt%	8.10	38.00	52.70	78.60	68.70	72.60	60.50
10. Product Solvent							
Carbon, Wt%	88.81	87.52	88.77	89.52	89.08	88.57	89.01
Hydrogen, Wt%	8.19	7.95	8.13	8.04	8.27	7.94	8.14
Sulfur, Wt%	2.26	2.23	2.26	2.25	2.36	2.24	2.35
Specific Gravity 60/60	1.050	1.056	1.058	1.056	1.054	1.056	1.055
Brookfield Visc., cp (1)	69.8	73.2	73.3	78.8	86.6	81.1	57.2
IR Ratio	0.20	0.22	0.21	0.22	0.21	0.21	0.21
11. Water							
Phenol, Wt%	0.80	3.81	4.47	5.03	6.20	7.30	6.90

(1) Brookfield Viscosity measured at approximately 23°C.

TABLE 2 - PART 1: MATERIAL BALANCE AND YIELD DATA

70-WEEK REACTIVITY SERIES TESTS AND NON-SOLVENT HYDROGENATION OF LIGNITE

	70-Week Reactivity Series Tests				Non-Solvent Hydrogenation	
Run No.	592	593	594	595	597	598
Test Conditions						
Time, hrs.	0.5	0.5	0.5	0.5	0.5	0.5
Avg. Temp., °F	752	752	752	752	751	752
Max. Press., psia	3275	3315	3315	3335	35 5	3 5
Gas Charged	CO/H_2					
Solvent	UNDCAO-75				None	
Lignite	73-2(B)				74-1(B)	
	N_2 Strg	N_2 Strg	Air Strg	H_2O Strg		
Material In, grams						
Coal	295.6	295.3	279.3	328.5	295.8	297.1
Water	28.7	28.9	44.5	1.8	18.5	18.0
Solvent	400.2	400.2	399.9	399.9	None	None
Gas	156.4	160.1	156.7	147.5	164.6	161.5
Total	880.9	884.5	880.4	877.7	478.9	476.6
Material Out, grams						
Filtrate	467.0	434.7	434.6	433.1	128.2(1)	142.3(1)
Filter Cake	41.8	69.5	62.4	82.7		
Residues	26.5	42.8	36.4	30.7	27.1	6.9
Gas	242.0	243.7	246.3	237.3	216.1	236.0
H_2O	63.4	74.7	67.8	74.7	92.6	84.2
Light Oil	9.5	2.2	2.1	3.7	2.3	4.3
Total	850.2	867.6	849.6	862.2	466.3	473.7
% Recovery	96.5	98.1	96.5	98.2	97.4	99.4
Yields 100% Recovery Basis						
Products, grams						
Light Oil	47.1	36.6	27.6	36.6	2.4	4.3
Solvent	381.3	372.9	367.1	374.1	---	---
Vac Btms	99.2	113.8	123.0	103.5	---	---
Total Liquid	527.6	523.3	517.7	514.2	68.8	107.2
Net Liquids	127.4	123.1	117.8	114.3	68.8	107.2
Net Gas	94.3	88.3	98.5	94.1	57.3	75.9
Net H_2O	-34.2	-23.7	-29.6	-23.5	-5.1	-14.8
Net Ash	+0.3	-2.3	-1.2	-3.9	0.4	-1.3
Unconverted Coal (MAF)	12.4	14.8	14.6	18.5	78.9	33.5
Yields Wt% MAF Coal						
Net Gas	47.1	44.1	49.2	47.2	28.6	37.9
Net Liquid	63.6	61.5	58.9	57.3	34.3	53.5
Unconverted	6.2	7.4	7.3	9.3	39.3	16.7
Net H_2O + Ash	-16.9	-13.0	-15.4	-13.8	-2.3	-8.1
Total Pct.	100.0	100.0	100.0	100.0	100.0	100.0
Filtrate Composition Wt%						
Light Oil	6.9	6.6	4.9	6.4	---	---
Solvent	70.7	71.6	71.2	73.3	---	---
Vacuum Btms	22.4	21.8	23.9	20.3	---	---
Net Liquid						
Net Light Oil	23.5	18.3	13.8	18.4	---	---
Net SRL	40.1	43.2	45.1	38.9	---	---
Solvent Recovery						
Wt%	92.0	91.4	88.6	91.9	---	---
Wt% (100% Rec. Basis)	95.3	93.2	91.8	93.6	---	---

(1) Unfiltered coal liquid.

TABLE 2 - PART 2: ANALYTICAL DATA

70-WEEK REACTIVITY SERIES TESTS AND NON-SOLVENT HYDROGENATION OF LIGNITE

Run No.	592	593	594	595	597	598
Test Conditions						
Time	0.5	0.5	0.5	0.5	0.5	0.5
Avg. Temp., °F	752	752	752	752	751	752
Max. Press., psia	3275	3315	3315	3335	3550	3340
Gas Charged	CO/H_2					
Solvent	UNDCAO-75					
Lignite	73-2(B)				74-1(B)	
	N_2 Strg	N_2 Strg	Air Strg	H_2O Strg		
Analytical Data						
1. Gas Analysis, Mol%						
H_2	42.8	47.3	42.2	44.9	43.5	46.2
CH_4	1.8	2.1	2.0	1.3	1.4	1.6
CO	18.0	25.8	22.9	22.9	34.0	26.2
C_2H_6	0.6	0.9	0.7	0.6	0.4	0.5
CO_2	36.3	23.3	31.7	29.9	20.4	25.3
H_2S	0.5	0.6	0.5	0.4	0.3	0.2
Gas Specific Gravity @ RT	0.8116	0.8092	0.8256	0.7859	0.704	0.726
2. Input Coal						
Volatile Matter, Wt%	35.74	35.76	34.93	28.71	29.69	32.34
Ash, Wt%	8.14	8.15	8.51	9.55	4.64	5.09
Moisture, Wt%	24.09	24.04	19.83	29.73	27.54	27.45
Carbon, Wt% (1)	63.95	64.29	63.25	62.53	67.20	65.37
Hydrogen, Wt% (1)	4.24	4.42	4.25	4.31	4.95	5.03
Sulfur, Wt% (1)	0.92	0.96	1.02	0.94	0.62	0.71
Nitrogen, Wt% (1)	---	---	---	---	0.79	0.86
Phenol, Wt%						
3. Input Solvent					None	None
Ash, Wt%	---	---	0.03	0.05	---	---
Carbon, Wt%	90.85	90.86	90.20	90.86	---	---
Hydrogen, Wt%	6.15	6.03	6.27	6.03	---	---
Sulfur, Wt%	0.46	0.52	0.53	0.55	---	---
Specific Gravity 60/60	1.113	1.113	1.114	1.120	---	---
Brookfield Visc., cp (2)	34.0	33.5	27.8	27.9	---	---
IR Ratio	1.78	1.69	1.77	1.81	---	---
4. Coal-Solvent Slurry						
Brookfield Visc., cp (2)	1485	2412	1322	2732	---	---
5. Cake					(3)	(3)
Ash, Wt%	24.44	27.95	31.95	30.02	8.90	9.19
Carbon, Wt%	64.68	61.61	56.14	59.50	77.06	77.96
Hydrogen, Wt%	4.20	3.93	3.50	4.11	5.51	5.87
Sulfur, Wt%	1.65	1.76	2.33	0.94	0.58	0.69
Nitrogen	---	---	---	---	1.05	1.16
Pyridine Sol., Wt%	82.77	74.41	71.22	70.00	45.73	75.47

(1) Calculated on a moisture-free basis

(2) Brookfield viscosity measured at approximately 23°C

(3) Unfiltered coal liquid

TABLE 2 - PART 2 CONT.: ANALYTICAL DATA

70-WEEK REACTIVITY SERIES TESTS AND NON-SOLVENT HYDROGENATION OF LIGNITE

Run No.	592	593	594	595	597	598
6. Filtrate						
Ash, Wt%	2.64	---	0.04	0.09	---	---
Carbon, Wt%	87.36	90.19	89.62	89.16	---	---
Hydrogen, Wt%	6.59	6.58	6.60	6.42	---	---
Sulfur, Wt%	0.43	0.45	0.29	0.30	---	---
Specific Gravity 60/60	1.103	1.129	1.128	1.126	---	---
Brookfield Visc cp (2)	1319	746	972	833	---	---
Blackness, abs./g. @ 550 mu	7.65	6.81	7.14	7.13	---	---
7. Vacuum Btms						
Ash, Wt%	11.49	0.98	0.42	0.96	---	---
Carbon, Wt%	78.09	86.51	87.88	89.29	---	---
Hydrogen, Wt%	5.10	5.96	5.85	5.64	---	---
Sulfur, Wt%	0.72	0.23	0.21	0.23	---	---
Melting Pt., °F	375	296	356	446	---	---
8. Residue						
Ash, Wt%	5.10	4.52	4.79	5.87	---	---
Carbon, Wt%	85.32	84.24	85.21	85.99	72.19	---
Hydrogen, Wt%	6.17	6.02	6.22	6.25	7.75	---
Sulfur, Wt%	0.85	0.67	0.45	0.50	0.53	---
Pyridine Sol., Wt%	99.32	95.87	94.68	97.47	---	---
9. Light Oil						
Carbon, Wt%	83.77	82.99	85.84	85.25	60.72	72.26
Hydrogen, Wt%	8.26	8.33	8.50	8.34	9.48	9.43
Sulfur, Wt%	0.51	0.33	0.68	0.48	---	0.16
10. Product Solvent						
Ash, Wt%	---	---	0.07	0.01	---	---
Carbon, Wt%	90.17	90.88	90.50	90.78	---	---
Hydrogen, Wt%	6.42	6.53	6.76	6.40	---	---
Sulfur, Wt%	0.41	0.40	0.47	0.44	---	---
Specific Gravity 60/60	1.111	1.109	1.108	1.107	---	---
Brookfield Visc., cp (2)	53.4	47.8	49.8	50.0	---	---
IR Ratio	0.90	0.91	0.85	0.83	---	---
11. Water						
Phenol, Wt%	---	---	---	---	---	---

(1) Calculated on a moisture-free basis
(2) Brookfield Viscosity measured at approximately 23°C
(3) Unfiltered coal liquid

Table 3

Hydrogenation of SRL in One-Liter and One-Gallon Autoclaves

Run	72	73(HDR-3)	74(HDR-4)	75
Hydrogen Uptake, psi	900	900	900	600
Output Gas Volume, cu. ft.	2.25	7.22	---	2.65
H_2O, grams	6.9	19.0	17.0	1.5
NH_3, grams	---	---	---	0.47

Distillation Fractions	Wt.gm	(%)	Wt.gm	(%)	Wt.gm	(%)	Wt.gm	(%)
1_A IBP-200°C/Atm	19.8	(26.4)	40.2	(14.1)	45.7	(16.1)	4.0	(5.3)
1_B IBP-87°C/1 Torr	7.6	(10.1)	36.8	(12.9)	36.9	(13.0)	3.5	(4.7)
2 87°-139°C/1 Torr	9.2	(12.3)	37.5	(13.1)	23.8	(8.4)	5.6	(7.5)
3 139-200°C/1 Torr	12.0	(16.0)	29.8	(10.4)	32.8	(11.5)	19.7	(26.2)
4 200-235°C/1 Torr	1.7	(1.4)	5.5	(1.9)	5.7	(2.0)	8.5	(11.3)
5 Vac. Bottoms	9.8	(12.2)	20.1	(7.2)	17.7	(6.2)	17.9	(23.8)
Overall Conversion		87.8		92.8		93.8		76.2

Gas Out								
H_2			13.4	47.8 (16.7)	12.4	49.48 (17.4)		
CH_4			10.5		7.68			
C_2H_6			12.9		17.4			
C_3H_8			24.4		24.4			
H_2S			0.9		0.9			

TABLE 4: HYDROGENATION OF SOLVENT REFINED LIGNITE

Run	Reaction Conditions	Distillation Fractions, gm (%)							
		1_A	1_B	2	3	4	Vacuum Bottoms	Total Conv.,%	NH_3 Produced, grams
76	75 g SRL, 150 ml Tetralin 2500 psi H_2, not stirred			21.1 (28.1)	13.8 (18.4)	1.5 (2.0)	30.2 (40.3)	59.7	0.78
77	No solvent, 1800 psi H_2 Conditioned in H_2, 4 days	10.1 (13.47)	3.2 (4.27)	6.4 (8.53)	13.4 (17.8)	4.3 (5.7)	9.5 (12.7)	87.3	0.75
78	75 g SRL, 150 ml Tetralin 2500 psi H_2, stirred	34.5	89.7	12.0 (16.0)	17.8 (23.7)	7.0 (9.3)	14.2 (14.9)	85.1	0.79
79	No Solvent, not conditioned Not exposed to air	11.7 (15.6)	7.10 (9.47)	8.5 (11.33)	15.8 (21.07)	4.8 (6.4)	8.8 (11.7)	88.3	0.85
80	SRC-Cu-88-18 No solvent	18.2 (24.27)	5.6 (7.5)	5.8 (7.7)	11.4 (15.2)	5.3 (7.0)	10.4 (13.9)	86.1	1.00

1_A	IBP - 200°C/Atm
1_B	IBP - 87°C/1 Torr
2	87° - 139°C/1 Torr
3	139 - 200°C/1 Torr
4	200 - 235°C/1 Torr

TABLE 5

NMR ANALYSES, HYDROGENATION EXPERIMENTS

Ar-H/Al-H Ratios of Distillation Fractions

	Distillation Fractions				
Run	1_A	1_B	2	3	4
72	0.143	0.283	0.316	0.480	0.640
74	0.121	0.232	0.302	0.525	0.620
75	0.143	9.516	0.537	0.829	0.882
76	0.669	0.180	3.340	0.216	2.340
77	0.049	0.382	0.377	0.700	0.861
78	0.289	0.482	1.410	1.420	1.490
79	0.085	0.232	0.252	0.392	0.647
80	0.081	0.251	0.258	0.397	0.600

Table 6

ECONOMIC SUMMARY
Standard SRL Process
(Mid - 1974 Prices)

Investment	M$ Invested
Total Plant Cost (insurance and tax basis)	196,320
Start-up Costs	5,890
Depreciation Subtotal	202,210
Return on Equity during Construction	24,501
Working Capital	10,000
Total Investment	236,711
Operating Costs[a]	
Fuel gas (5,533 MM Btu/hour @ 60¢/MM Btu)	26,293
Lignite (1,474 tons/hour @ 2.25/ton)	26,267
Operating Labor (200 men @ $5/hour)	2,080
Maintenance Labor and Material	6,933
Supervision and Technical Service	728
Other Direct Production Costs	1,623
Total Direct Production Costs	63,924
Fixed Charges	13,144
Plant Overhead and General Expense	3,567
Total Operating Costs	80,635
By-Product Credits	
Electricity (61,035 KWH/hour @ 0.5¢/KWH)	2,417
Sulfur (220 LT/day @ $15/LT).	1,089
Total By-Product Credits	3,506
Total Operating Costs After Credits	77,129

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[a] Start-up, operating costs of $13,742 million incurred during the last year of construction are not shown.

Table 7

ECONOMIC SUMMARY
SRL with Lurgi
(Mid-1974 Prices)

Investment	M$ Invested
Total Plant Cost	196,322
Start-up Costs	5,890
Depreciation Subtotal	202,212
Return on Equity during Construction	24,501
Working Capital	10,000
Total Investment	236,713
Operating Costs[a]	
Lignite (1,617 tons/hour @ $2.25/ton)	28,815
Operating Labor (221 men at $5/hour)	2,298
Maintenance Labor and Material	6,937
Supervision and Technical Service	761
Other Direct Production Costs	1,623
Total Direct Production Costs	40,434
Fixed Charges	13,144
Plant Overhead and General Expense	3,729
Total Operating Costs	57,307
By-Product Credits	
Electricity (32,955 KWH/hour @ .05¢/KWH)	1,305
Sulfur (224 LT/day @ $15/LT)	1,108
Total By-Product Credits	2,413
Total Operating Costs After Credits	54,844

---

[a] Start-up, Operating Costs of $13,743 million incurred during last year of construction are not shown.

Table 8

PRODUCT SELLING PRICE
SSRL and SRLL PROCESSES
(Mid - 1974 Prices)

Item	SSRL		SRLL	
DCF Rate (%)	$/ton	¢/MM Btu	$/ton	¢/MM Btu
10	26.216	76.2	20.188	58.5
12	28.508	82.9	22.452	65.1
14	30.958	90.0	24.872	72.1
Coal Cost ($/ton)				
0	21.694	63.1	15.068	43.7
2.025	27.826	80.9	21.712	62.9
2.25	28.508	82.9	22.452	65.1
2.475	29.190	84.9	23.190	67.2
Labor Cost (1.00 ± .10)[a]				
.90	28.272	82.2	22.206	64.4
1.00	28.508	82.9	22.452	65.1
1.10	28.746	83.6	22.696	65.8

---

[a] Base figures are $9.148 million for SSRL and $9.563 million for SRLL respectively. Labor cost includes direct, maintenance, and indirect labor along with supervisory and technical services and payroll overhead.

Table 9

ECONOMIC SUMMARY
SLAGGING LURGI GASIFICATION
(Mid-1974 Prices)

Investment	$M Invested
Total Plant Cost	20,769
Start-up Cost	623
Depreciation Subtotal	21,392
Return on Investment during Construction	2,592
Working Capital	1,000
Total Investment	24,984
Operating Costs[a]	
Lignite (143 tons/hour @ $2.25/ton)	2,548
Operating Labor (20 men @ $5/hour)	208
Maintenance Labor and Material	721
Supervision and Technical Services	74
Other Direct Production Costs	295
Total Direct Production Costs	3,846
Fixed Charges	1,390
Plant Overhead and General Expense	365
Total Operating Costs	5,601
By-Product Credits	
Steam (11,300 lbs. of 50psig/hr. @ $.5/MM Btu)	45
Sulfur (4LT/day @ $15/LT)	20
Total By-Product Credits	65
Total Operating Costs After Credits	5,536

---

[a] 1st year start-costs (7% Total Plant Costs) expensed in year 0 are not shown.

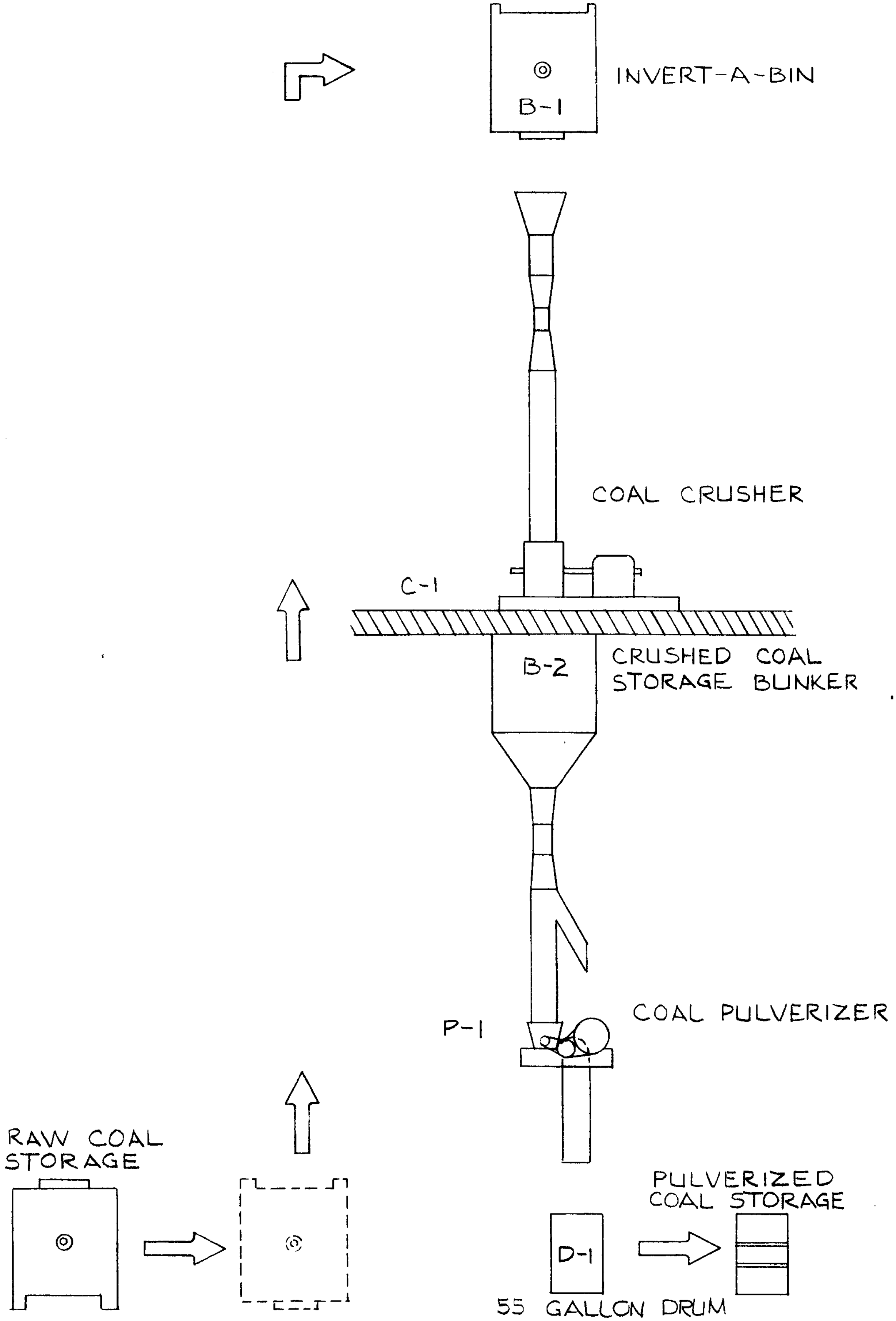
Table 10

PRODUCT SELLING PRICE
SLAGGING LURGI GASIFICATION
(Mid - 1974 Prices)

Item	¢/MM Btu
DCF Rate (%)	
10	61.3
12	68.4
14	76.0
Coal Cost ($/ton)	
0	49.0
2.025	66.5
2.25	68.4
2.475	70.4
Labor Cost (1.00 ± .10)[a]	
.90	67.7
1.00	68.4
1.10	69.2

---

[a] Base figure is $935.7 thousand and includes direct, maintenance, and indirect labor along with supervisory and technical services and payroll overhead.



COAL CRUSHING AND GRINDING AREA 01A

FIGURE 1

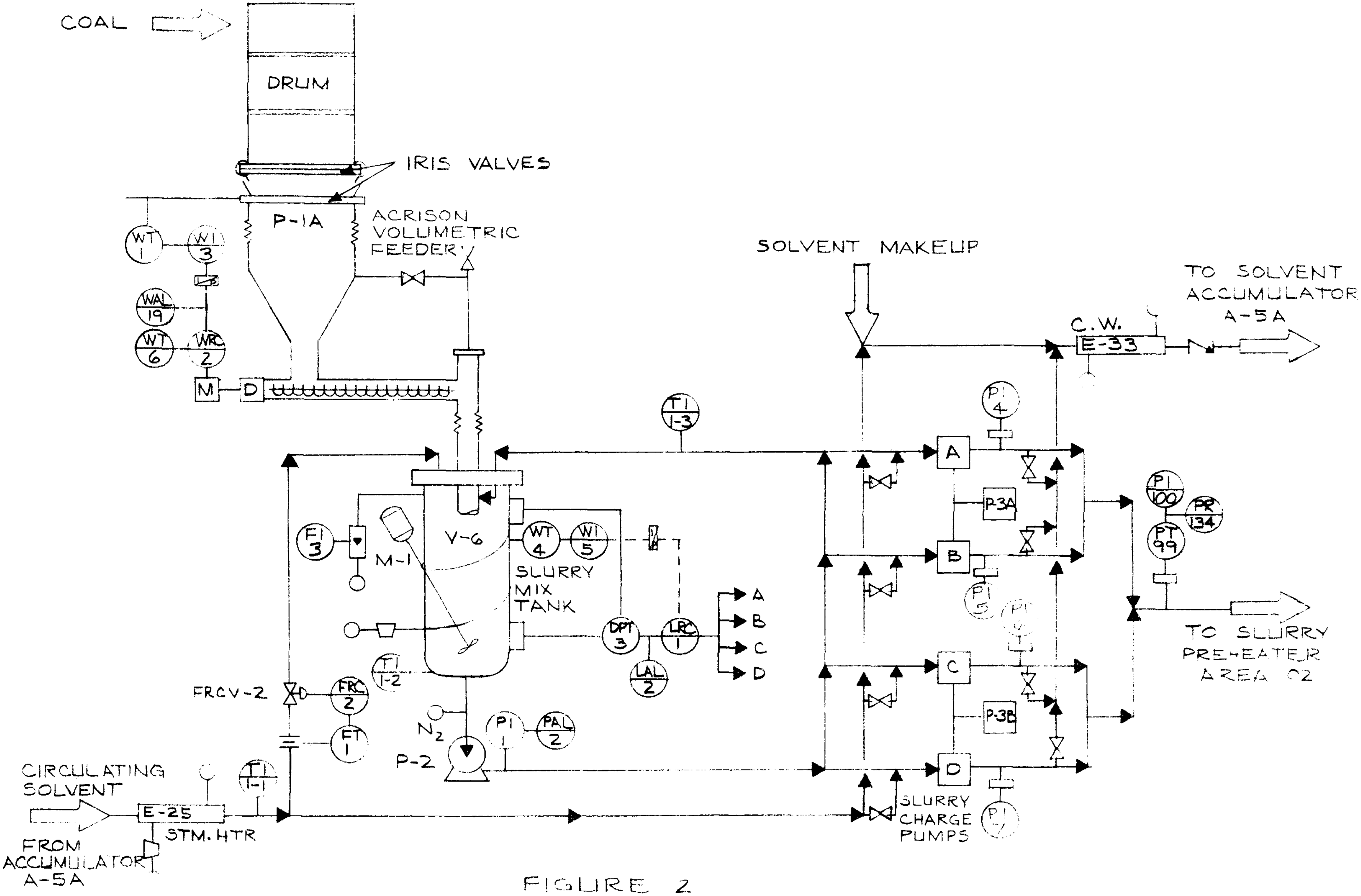


FIGURE 2

SLURRY PREPARATION & PUMPING
AREA 01B

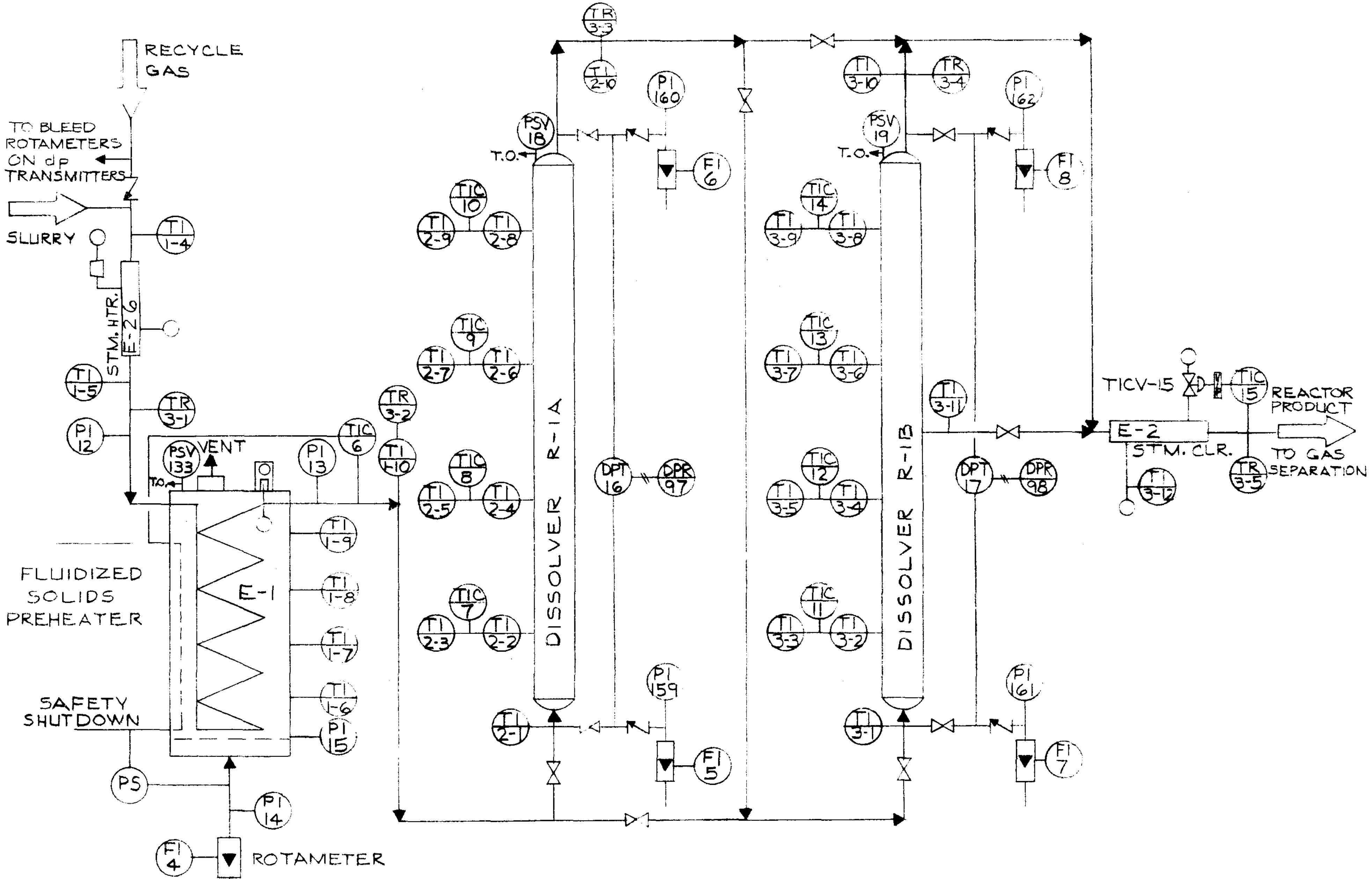


FIGURE 3

SLURRY PREHEATING & DISSOLVING
AREA 02A

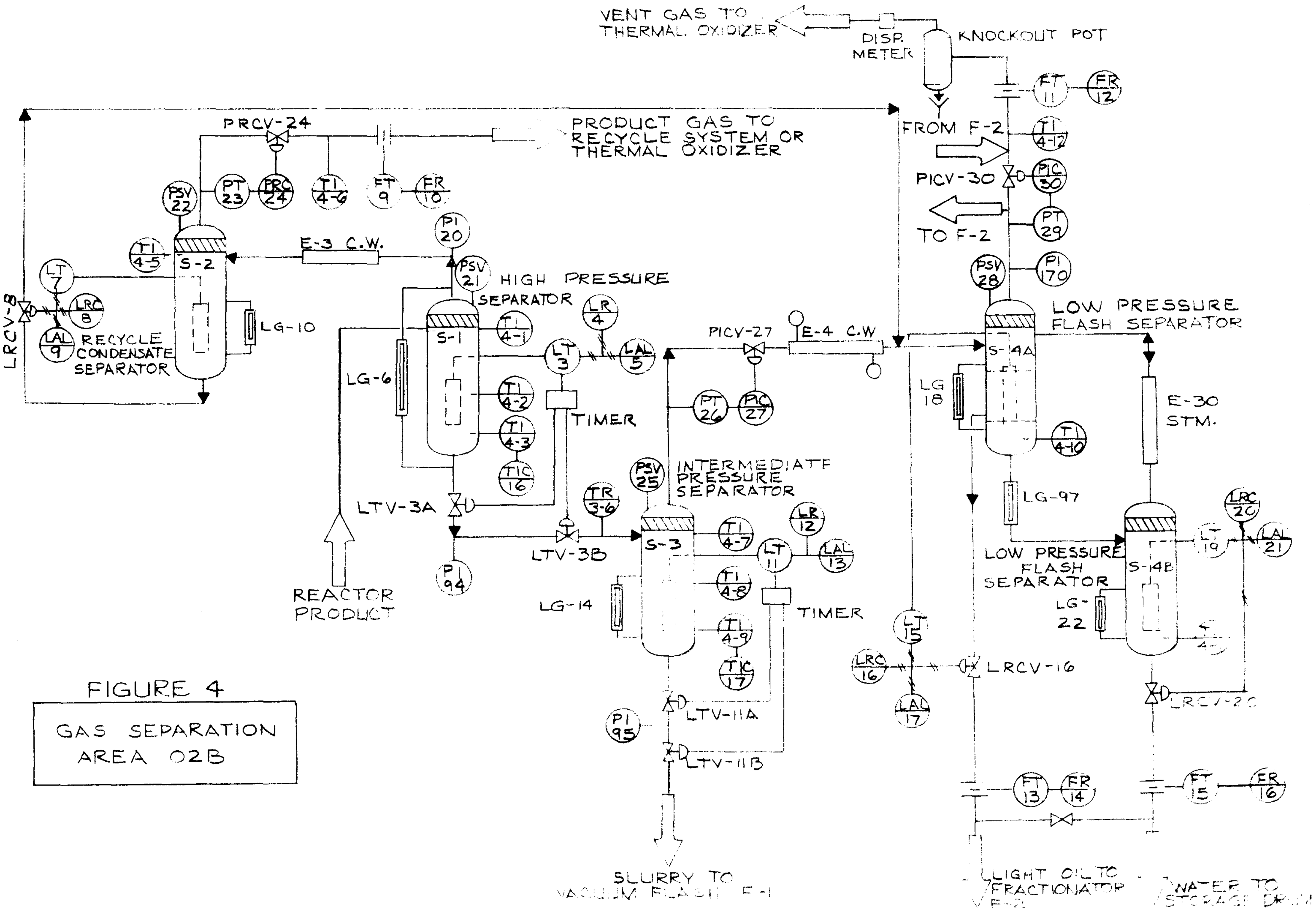


FIGURE 4

GAS SEPARATION AREA 02B

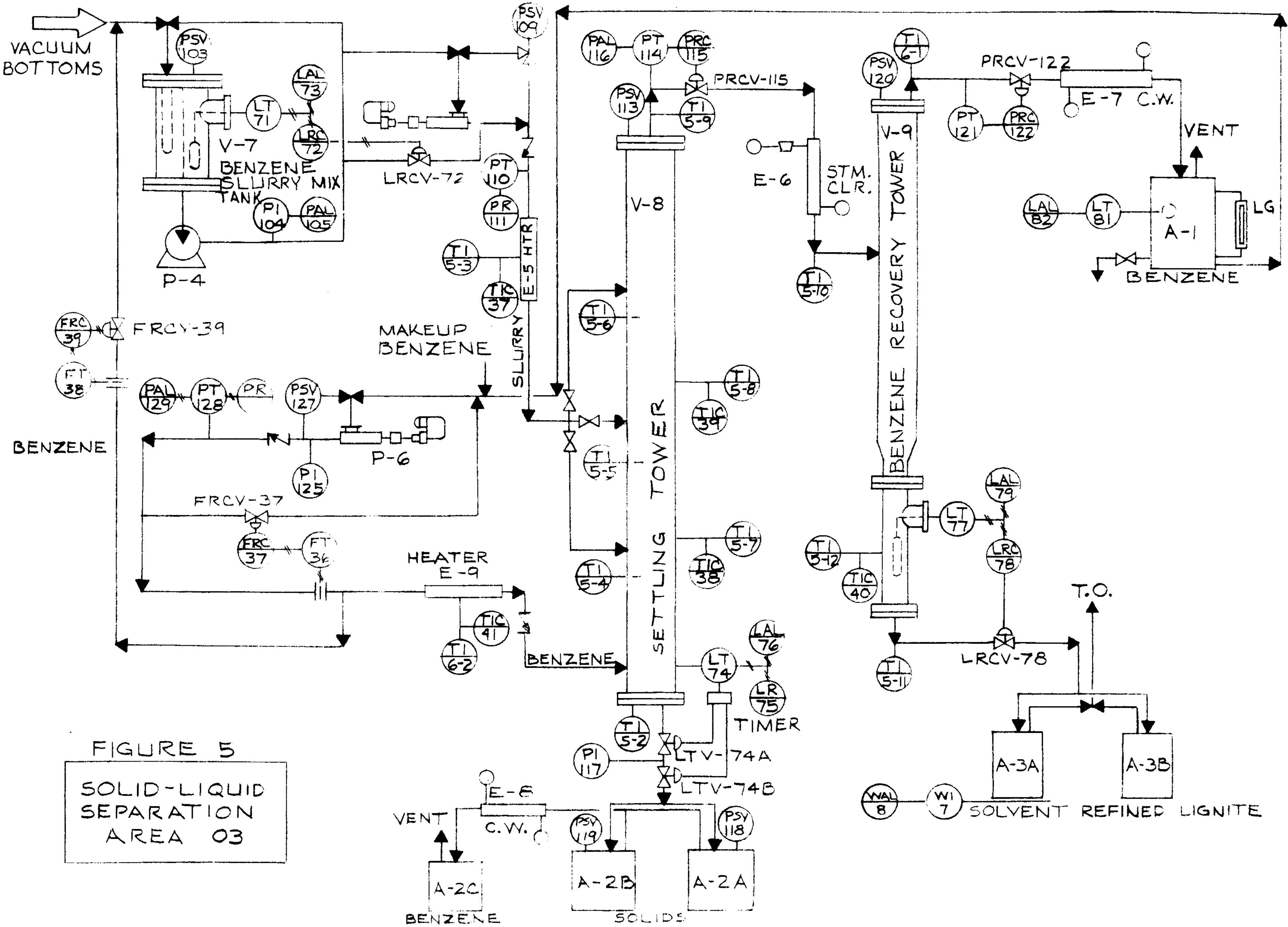
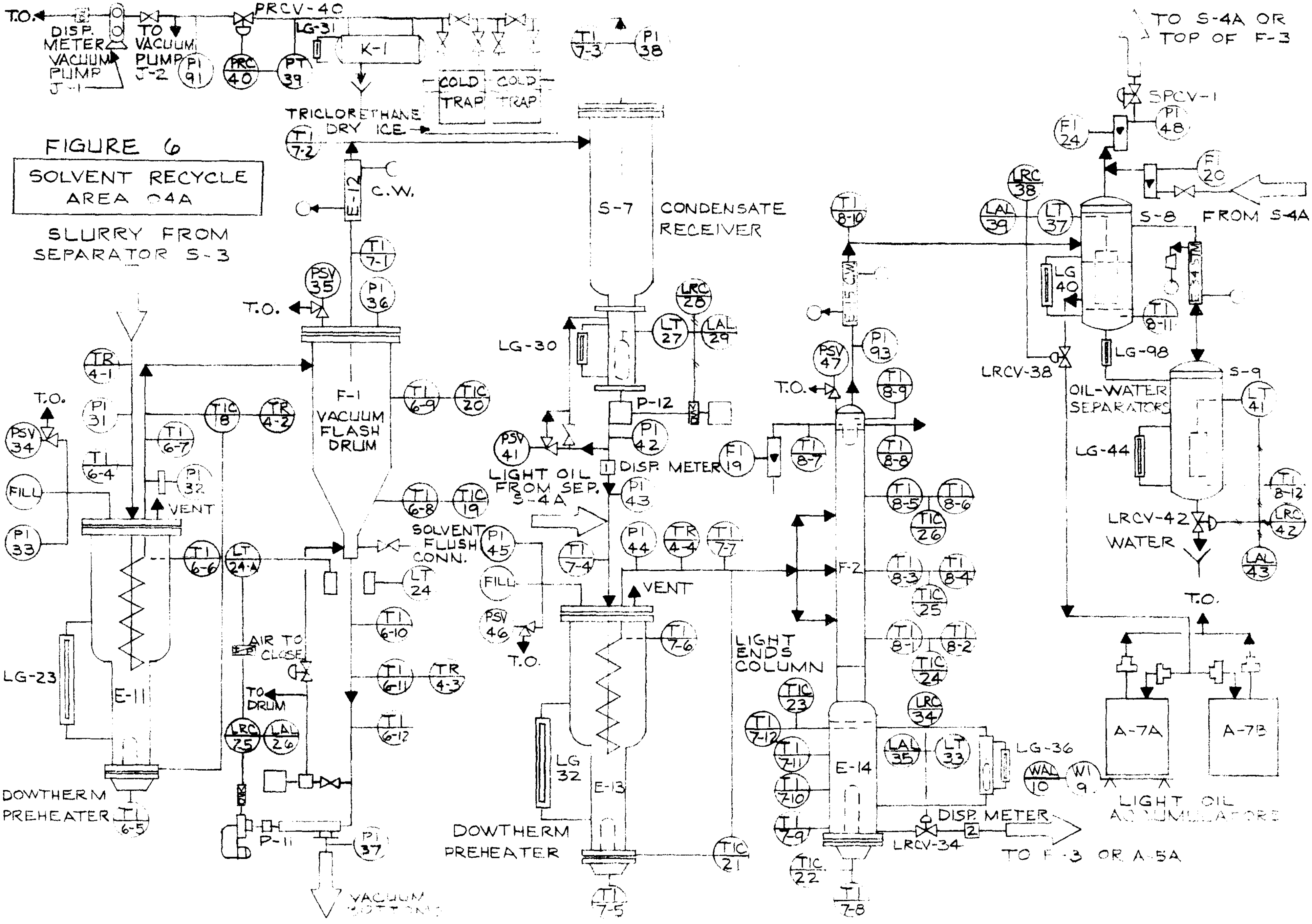


FIGURE 5

SOLID-LIQUID SEPARATION AREA 03

FIGURE 6

SOLVENT RECYCLE AREA 04A



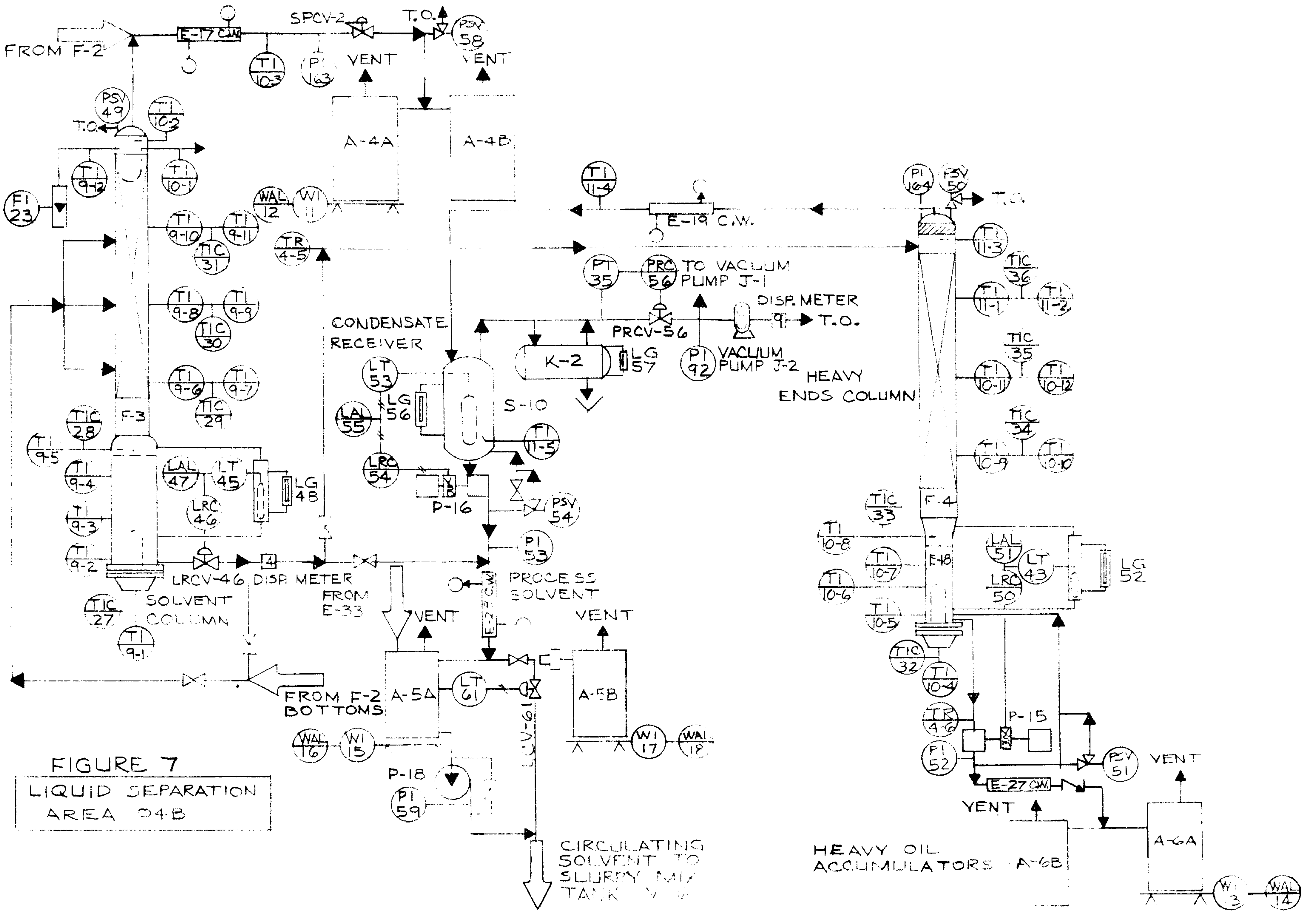


FIGURE 7

LIQUID SEPARATION
AREA 94-B

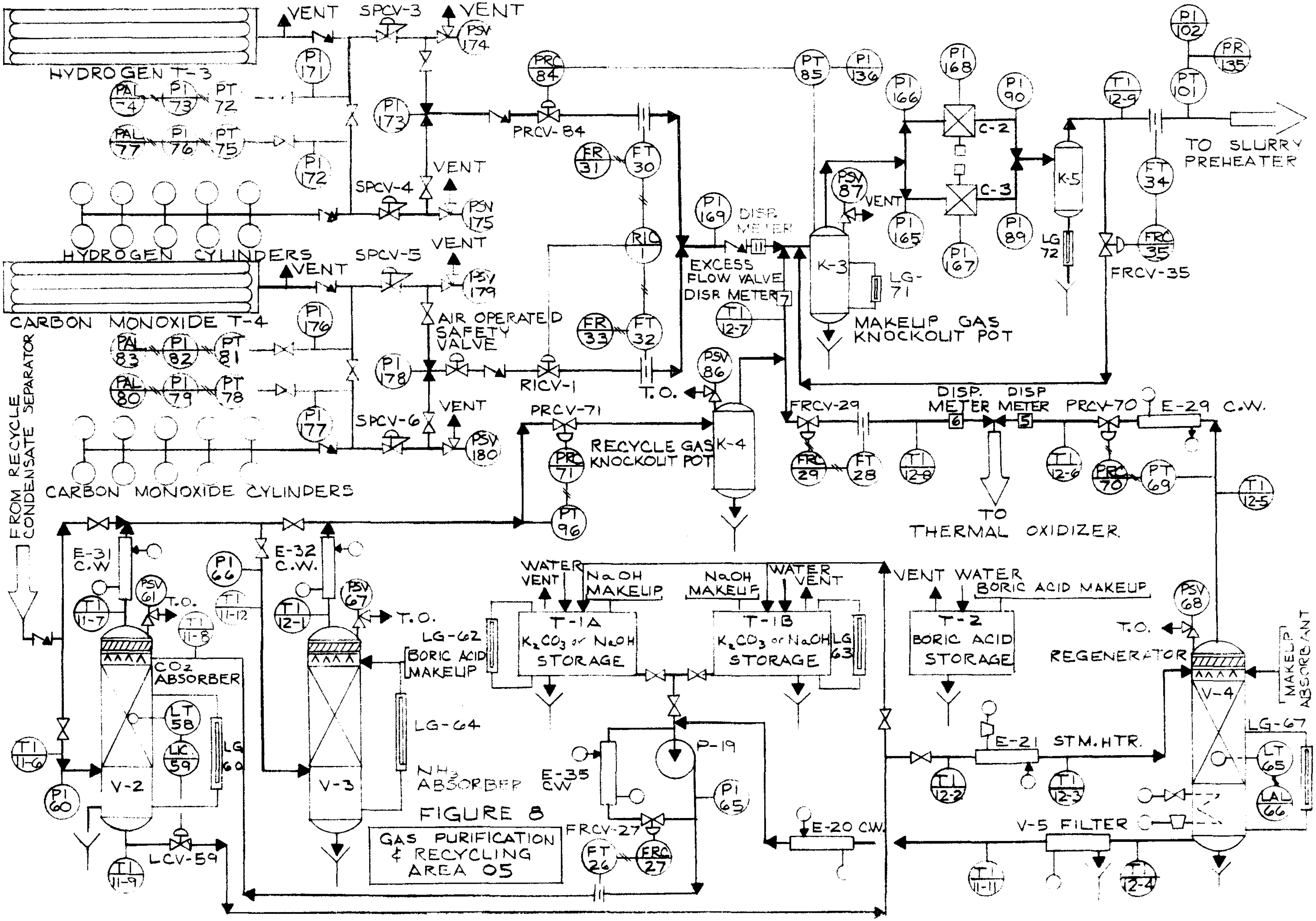


FIGURE 8

GAS PURIFICATION & RECYCLING AREA 05

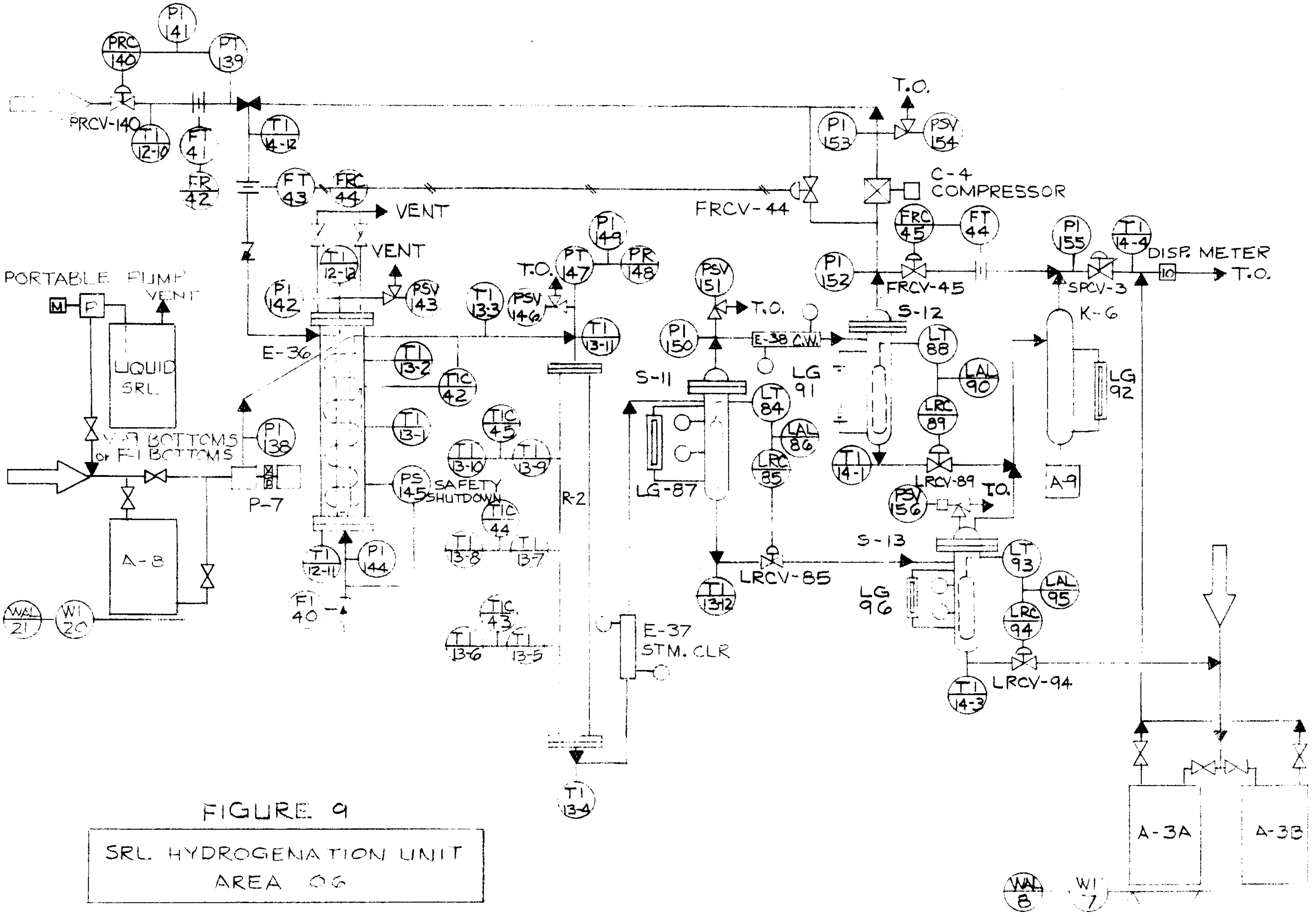


FIGURE 9

SRL HYDROGENATION UNIT
AREA 06