LIST OF ACID GAS REMOVAL PROCESSES

TYPE	SOLVENT	MODE
AD = Adsorption	C = Chemical Solvent	S = Selective
AB = Absorption	P = Physical Solvent	NS = Non-selective
CD = Cryogenic Distillation		

	NAME OF PROCESS	TYPE	SOLVENT	MODE	MAJOR CONTAMINANTS REMOVED
1	Activated Carbon	AD	C	NS	H_2S,oil
2	ADIP	AB	C	NS	H_2S,CO_2
3	Alkazid	AB	C	S,NS	H_2S,CO_2
4	Amisol	AB	C/P	S,NS	H_2S,CO_2
5	Benfield	AB	C	NS	H_2S,CO_2
6	Catacarb	AB	C	NS	H_2S,CO_2
7	Chemsweet	AD	C	S	H_2S
8	CNG	AB	P	S	H_2S,CO_2
9	Estasolvan	AB	P	S,NS	H_2S,CO_2,oil
10	Flexsorb SE	AB	C	S	H_2S,CO_2
11	Fluor Econamine	AB	C	NS	H_2S,CO_2
12	Fluor Solvent	AB	P	NS	H_2S,CO_2,oil
13	Giammarco-Vetrocoke	AB	C	S	H_2S,CO_2
14	MEA	AB	C	NS	H_2S,CO_2
15	MDEA	AB	C	S,NS	H_2S,CO_2
16	Molecular Sieves	AD	P	S	H_2S
17	Purisol	AB	P	S,NS	H_2S,CO_2
18	Rectisol	AB	P	S,NS	H_2S,CO_2
19	Ryan Holmes	CD	CD	S	H_2S,CO_2,C_2+
20	Seaboard	AB	C	S	H_2S
21	Selexol	AB	P	S,NS	H_2S,CO_2,oil
22	Sepasolv MPE	AB	P	S,NS	H_2S,CO_2,oil
23	SNPA - DEA	AB	C	NS	H_2S,CO_2
24	Stretford	AB	C	S	H_2S
25	Sulfiban	AB	C	NS	H_2S,CO_2
26	Sulfinol	AB	C/P	NS	H_2S,CO_2
27	Tripotassium Phosphate	AB	C	S	H_2S
28	Vacuum Carbonate	AB	C	S	H_2S
29	Zinc Oxide	AD	C	S	H_2S

ACTIVATED CARBON

CALGON CORPORATION, PITTSBURGH, PENNSYLVANIA
UNITED CATALYSTS, INC., LOUISVILLE, KENTUCKY

- Adsorption process for removal of H_2S and other low molecular weight sulfur compounds from industrial gases, using beds of carbon impregnated with 5-10% of metallic oxides.
- No pressure limitation; temperature limit for the feed gas is 150°F. Regeneration of bed at 450-500°F.
- Process is considered most applicable for desulfurizing gas from 30 ppm H_2S to 0.2 ppm.
- More than 60 commercial plants using the process are in operation in U.S.A.

ADIP PROCESS

SHELL DEVELOPMENT COMPANY, HOUSTON, TEXAS

- Process uses an aqueous solution of di-isopropanolamine (DIPA) to remove H_2S from industrial gases. Process is also used for treating LPG and NGL.
- Absorber Pressure: 255 PSIG
  Temperature: 100 to 140°F

  Regenerator Pressure: Near Atmospheric
  Temperature: 250 to 275°F
- Process can reduce H_2S content in natural gas to less than 5 ppmv and in synthesis gas to less than 100 ppmv.
- More than 150 plants have been built worldwide.

ALKAZID

DAVY McKEE AG, WEST GERMANY

- For the selective absorption of H_2S and for the simultaneous removal of H_2S and CO_2 at atmospheric or higher pressures. Two different solvents: Alkazid "M" and Alkazid "DIK".

- Absorption:
  - Pressure: 0 to 1000 PSIG
  - Temperature: Ambient or below ambient

- Purification: At low pressure, removal to 85 ppm H_2S with CO_2 in feed gas; to 320 ppm H_2S without CO_2 in feed gas.

- More than 100 plants in operation; none in U.S.

- Suitable for synthesis gas, water gas, natural gas and hydrogen.

AMISOL

LURGI KOHLE UND MINERALOELTECHNICK GMBH., W. GERMANY

- Process for selective desulfurization or complete removal of CO_2, H_2S, COS and other organic sulfur compounds from synthesis gas, using a mixture of alkanolamine (MEA, DEA) with methanol as a solvent.

- Absorption:
  - Temperature: Ambient
  - Pressure: Above 200 PSIA

  Regeneration:
  - Temperature: Below 212°F
  - Pressure: Little more than atmospheric

- One plant onstream and two under construction with a total throughput of 73 MM SCFD.

BENFIELD

BENFIELD CORPORATION, PITTSBURGH, PA

- For the removal of H_2S and CO_2 from feed gas mixture; provides H_2S-rich stream suitable for recovery of elemental sulfur. Hot potassium carbonate solution with Benfield additives is used for absorption.
- Absorber Pressure: 100 to 2000 PSIG
  Temperature: Ambient to 400°F
- Purification Levels: H_2S: From 10% in feed to ppm level in product gas
  CO_2: From 25-45% in feed to ppm level in product gas
- More than 500 units have been licensed, including over 210 for purifying SNG, reformed and partial oxidation gases.
- Removes most of the COS, HCN and CS_2; is not guaranteed for removal of thiophene.

CATACARB

EICKMEYER AND ASSOCIATES, OVERLAND PARK, KS

- For removal of CO_2, H_2S and COS from gas streams. Raw gas is scrubbed in a packed or trayed absorber with a potassium salt solution containing a very active, stable and non-toxic catalyst.
- Absorption Pressure: 40 to 1000 PSIG
  Temperature: 100 to 260°F
- Process can reduce H_2S to about 4 ppm and CO_2 to as low as 50 ppm.
- About 100 plants are in operation.
- Sour acid gases require further processing to recover sulfur.

CHEMSWEET

C-E NATCO, TULSA

- A chemical adsorption system used for the continuous removal of H_2S and other sulfurous gases from natural gas.

- Chemsweet powder is a mixture of zinc oxide and zinc acetate. The sour gas is bubbled through a water slurry of Chemsweet powder.

- Hydrogen sulfide, mercaptans, and carbonyl sulfide react instantaneously with zinc acetate to form insoluble zinc sulfide and zinc mercaptide. The zinc oxide replenishes the zinc acetate as it is consumed.

- Operating pressure: 75 to 1400 psig.

  Operating temperature: Ambient

  The CO_2 content of the gas does not have an impact on the process chemistry because the pH of the Chemsweet slurry is too high to produce zinc carbonate.

- Over 60 commercial installations in the U. S.

CNG

CONSOLIDATED NATURAL GAS RESERCH COMPANY

- The CNG process is a cryogenic physical absorption process that removes acid gases (CO_2, H_2S, COS and other trace contaminants) from medium to high pressure gas streams containing substantial amounts of CO_2, such as synthesis gas produced by coal gasification or heavy oil partial oxidation, and natural gas in enhanced oil recovery projects.

- The process has three novel features:

  (i) Absorption of sulfur-compounds and trace contaminants with pure liquid CO_2.

  (ii) Regeneration of pure CO_2 with simultaneous concentration of H_2S and trace contaminants by triple-point crystallization.

  (iii) Absorption of CO_2 with a slurry of organic liquid containing CO_2.

- The sulfur absorber operates at -67°F, slightly above the triple point of CO_2. The feed gas pressure can range from 300 to 1500 psia. The triple-point crystallizer operates at pressures and temperatures near the triple point of CO_2. The CO_2 absorber operates at temperature range between -67°F and -112°F.

- The process is currently in advanced stages of development at the bench scale level. Future plans call for a 2-3 ton/day PDU.

ESTASOLVAN

INSTITUT FRANCAIS DU PETROLE, FRANCE AND FRIEDRICH UHDE, GMBH, W. GERMANY

- Process uses tributylphosphate as solvent and incorporates simultaneous absorption of acid gases and LPG in the solvent, followed by separation of the absorbed components by fractional distillation.
- Solvents have high capacity for acid gases and good selectivity for H_2S with respect to CO_2.
- Demonstrated in two pilot plants, but as yet no commercial application.

FLEXSORB SE

EXXON RESEARCH AND ENGINEERING COMPANY, FLORHAM PARK, NEW JERSEY

- A chemical absorption system that uses a hindered amine solution that can selectively remove H_2S.
- The process can be applied to sour low BTU fuel gas from Flexicoking and Claus Plant Tail Gas.
- The process operates under conditions similar to MDEA systems and uses the same equipment. However, it requires 41% of the MDEA circulation rate and 51% of its energy for regeneration.
- The process has been commercially tested at an Exxon refinery in a Claus Tail Gas clean-up unit, which had been designed to operate using MDEA solution.

FLUOR ECONAMINE

FLUOR ENGINEERS AND CONSTRUCTORS, INC., IRVINE, CA

- Diglycolamine (DGA) is the solvent used to absorb H_2S and CO_2 from sour gas.
- Absorber:

Temperature:	90-220°F
Pressure:	About 900 PSIG

  Regenerator:

Temperature:	250-280°F
Pressure:	5-15 PSIG

- Process can be designed to reduce H_2S concentration to 4 ppm and CO_2 to 50 ppm. DGA also removes COS.
- More than 30 plants processing six billion SCFD of natural gas are in operation.

FLUOR SOLVENT

FLUOR ENGINEERS AND CONSTRUCTORS, INC. IRVINE, CA

- o Process uses propylene carbonate ($C_3H_6CO_3$) for physical absorption of H_2S, CO_2 and other organic sulfur compounds. Desorption of acid gases is carried out by reducing solvent pressure in several stages. Economics favor high absorber pressure.
- o Pressure: 850 - 1000 PSIG
- o Reduces H_2S level to less than 4 ppmv and the CO_2 level to less than 0.3% vol %.
- o A total of 9 plants (5 on natural gas, 1 in hydrogen production, and 3 in ammonia production) are in operation.

GIAMMARCO - VETROCOKE

GIOMMARCO-VETROCOKE OF MARGHERA, ITALY

- o The process is used for the removal of H_2S by scrubbing the sour gas with a solution of alkali arsenates and arsenites; sodium carbonate is the alkali usually applied.
- o Pressure: 1100 PSIG in the absorber.
- o Reduces H_2S level to 0.15 ppm or less.
- o Approximately 50 plants have been built.

MEA

ALL ENGINEERING COMPANIES IN THE GAS PURIFICATION FIELD

- Monoethanolamine (MEA) has been the solvent for many years to remove (non-selectively) H_2S and CO_2 from natural and synthesis gases.

- Absorber
  Temperature: 110-160°F
  Pressure: Not sensitive

- Regenerator
  Temperature: 205 - 250°F
  Pressure: Near atmospheric

- MEA can reduce H_2S content to less than 1 ppm and CO_2 to less than 0.1% (V).

- Development of "Amine Guard" inhibitors by Union Carbide made the MEA process more attractive by reducing corrosion and lowering energy requirements.

MDEA

ALL ENGINEERING COMPANIES IN THE GAS PURIFICATION FIELD

- Process uses an aqueous solution of methyl-diethanolamine (MDEA) to remove H_2S by chemical absorption from industrial gases.

- Absorber
  Temperature: 80-125°F
  Pressure: 0-1000 PSIG

  Stripper
  Temperature: 240-250°F
  Pressure: 7-10 PSIG

- H_2S can be reduced to 4 ppm at low pressures. Process can not remove Mercaptans.

- At least 20 units in the U.S. and several in Europe are operating.

- Several companies have developed their own proprietary corrosion inhibitors and offer different versions of an activated MDEA systems. Most improvements have been made in two areas - improving H_2S selectivity and reducing energy consumption.

The following gives the various licensors of MDEA process:

Dow Chemical U.S.A:	Selectamine Process Gas/Spec FT-1 Technology
Union Carbide:	H-S Process
BASF:	Activated MDEA Process
Shell Oil Co.:	SAGA Process

MOLECULAR SIEVE

UNION CARBIDE CORPORATION, TARRYTOWN, NEW YORK

- Process to dehydrate and to remove carbon dioxide and sulfur compounds from natural gas, LPG and NGL. Process involves two or more fixed bed adsorbers and regeneration facilities.
- Normal Adsorption
  Pressure: 200 to 1200 PSIG
  Temperature: 60 to 120°F
- Sour natural gas can be desulfurized to 4 ppmv. CO_2 removal with molecular sieves is most attractive when the product must have a very low CO_2 content and the feed has less than 1.5% CO_2.
- Sieve life ranges from 2 to 10 years for desulfurization and carbon dioxide removal, and from two to six years for dehydration.
- More than 60 LNG plants use molecular sieves for drying and purification.

PURISOL

LURGI KOHLE AND MINERALOELTECHNICK GMBH, FRANKFURT, W. GERMANY

- A physical absorption process that offers high solubility at high acid gas partial pressures. The process uses n-methyl-2-pyrolidine (NMP) as the solvent to absorb CO_2, H_2S and other acid gases. The absorption is reversed by reducing the pressure.
- Absorber
  Pressure: Low to 1000 PSIG
  Temperature: 80-105°F
- Typically reduces H_2S to 4 ppmv and CO_2 to 2-3 vol%.
- Five plants are in operation, processing about 440 MM SCFD of feed gas.
- Process is typically used for removal of acid gases from natural gas, hydrogen and synthesis gas.

RECTISOL

LURGI KOHLE UND MINERALOELTECHNICK GMBH, FRANKFURT, W. GERMANY

- Process uses methanol for removal of acid gases by physical absorption at relatively low temperatures and high pressures. Acid gases are separated from the solvent by decreasing the pressure or by stripping. High solubility of H_2S in the solvent allows selective removal of H_2S.

- Pressure: 300 to 2000 PSIG
  Temperature: -100 to 0°F

- Purification Levels - H_2S and COS: 1 ppm
  CO_2: 1 ppm

- More than 37 plants with a total capacity of 6 billion SCFD are in operation.

- Used mainly in coal or oil gasification processes that produce fuel gas, synthesis gas for ammonia, methanol or SNG.

RYAN-HOLMES

KOCH PROCESS SYSTEMS, INC.

- The Ryan-Holmes Processes is a cryogenic distillation process used for separating acid gases (such as CO_2 and sulfur compounds) from natural gas streams. An NGL-based additive is introduced to a near the top of the distillation column to enhance the distillation.

- The distillation columns operate between 350 to 650 psia. The operating temperatures can range from -120°F to 40°F. Acid gas contents (mainly CO_2) in the feed can range from 30% to 90%.

- Koch Process Systems has licensed four applications of the Ryan/Holmes technology in enhanced oil recovery projects. These processes have not yet been applied, or proven to process synthesis gases from coal gasification units.

SEABOARD

KOPPERS COMPANY, INC.

- o Process is based upon the absorption of hydrogen sulfide by a dilute sodium carbonate solution and regeneration by air stripping.

- o Process is capable of removing 85 to 95% of the H_2S in the feed gas in a single stage.

- o Main drawbacks are:
  - Occurrence of side reactions caused by oxygen in the air
  - The disposal of foul air containing H_2S.

SELEXOL

NORTON COMPANY, STOW, OHIO

- o The solvent of the Selexol process is a dimethyl ether of polyethyleneglycol, which has a high physical absorption capacity for sulfur-based compounds including H_2S, COS and mercaptans, as well as for CO_2. The solvent is regenerated by stripping with steam or inert gas.

- o Pressure: 500 to 1000 PSIG
  Temperature: 50 to 200°F.

- o Purification Levels - H_2S, COS and mercaptans: 1 ppm. CO_2 can be retained or reduced to any required level.

- o 23 plants are with a total capacity of 1240 MM SCFD gas are in operation.

- o Highly selective for sulfur compounds. The solvent is physically and chemically stable, and is not degraded by impurities.

SEPASOLV MPE

BASF AKTIENGESELLSCHAFT

- o Sepasolv MPE is a physical solvent used for the removal of acid gases (CO_2, H_2S, COS and mercaptans) from natural gas and synthesis gas produced via coal gasification or heavy oil partial oxidtion.

- o Sepasolv MPE is a mixture of oligoethylene glycol and methyl isopropyl ethers. The Sepasolv MPE and the Selexol solvents are very similar in chemical structure and physical properties. Thus, their application range is almost identical. The solutions can be exchanged in most plants without any equipment modifications.

- o There are four operating plants, three of which are for the selective removal of H_2S from CO_2 containing natural gases and one is for the combined removal of H_2S and CO_2 from natural gas. Two of these plants had previously used Selexol.

SNPA-DEA

RALPH M. PARSONS CO., PASADENA, CALIFORNIA

- Diethanolamine is the solvent in this process for treating high pressure natural gases with high concentrations of acidic components, especially COS and CS_2. (SNPA stands for Societe National des Petroles d'Aquitaine).
- Absorber Pressure: 600 to 1100 PSIG
- The process can reduce H_2S to less than 3 ppm and CO_2 to 500 ppm in natural gas. COS and CS_2 do not deteriorate the solution.
- The process has been used to treat over 5 billion SCFD of sour gas.
- The solvent is mildly corrosive.

STRETFORD

PEABODY PROCESS SYSTEMS, STAMFORD, CONNECTICUT

- Typical aqueous solvent contains sodium carbonate, anthraquinone - disulfonic acid (ADA) and vanadium as sodium metavanadate. H_2S is removed by chemical absorption.
- Absorber
  - Temperature: 90 to 120°F
  - Pressure: Atmospheric to 100 PSIG
- H_2S can be reduced to 1 ppm. Process does not remove any significant amount of CO_2. Complete removal of HCN can be obtained. COS and CS_2 are not reduced to a significant degree.
- Main problem connected to process is the treatment and disposal of effluent streams containing sodium thiocyanate and sodium thiosulfate (from HCN).
- Over 50 plants are currently in operation worldwide.

SULFIBAN

BLACK, SIVALLS AND BRYSON, INC., HOUSTON, TEXAS

- Process uses an aqueous solution of monoethanolamine (MEA) with proprietary inhibitors to remove H_2S from industrial gases.

- Absorber
    Temperature: 100°F
    Pressure: 0 - 1000 PSIG

  Stripper
    Temperature: 200-250°F
    Pressure: 7-10 PSIG

- Process can reduce H_2S to about 1 ppmv, or remove more than 98% of H_2S in feed. Does not remove hydrocarbons from the feed gas. Solvent degradation occurs if COS and HCN are present in the feed gas.

- Plants to process more than 1.75 MM SCFD of sour gas have been built.

SULFINOL

SHELL DEVELOPMENT COMPANY, HOUSTON, TEXAS

- Process removes H_2S, CO_2, COS and organic sulfur from natural gas and synthesis gas. The solvent consists of an alkanolamine, water and an inert organic compound, tetrahydrothiophene dioxide ("Sulfolane").

- Pressure in the absorber may vary from slightly above atmospheric to 1400 PSIG. High pressure operation favors physical absorption.

- Reduces H_2S level to 1 ppmv and CO_2 level to less than 50 ppmv.

- Over 140 units are in operation or under construction, most of them for natural gas treating. Single location capacities vary from 15 to 500 MM SCFD.

TRIPOTASSIUM PHOSPHATE

SHELL DEVELOPMENT COMPANY, HOUSTON, TEXAS

- Process uses a tripotassium phosphate solution to remove H_2S. Principal advantages are non-volatility of the active component in the solution, insolubility in hydrocarbon liquids and non-reactivity with COS and other trace impurities.
- Suitable for high temperature applications.
- Process has been displaced by ethanolamine processes.

VACUUM CARBONATE

KOPPERS COMPANY, INC.

- Process is an outgrowth of the Seaboard process. Use of vacuum distillation for regeneration of the alkali-carbonate solution reduced steam requirement to about one-sixth.
- Process has been primarily applied to remove H_2S from coke oven gases.
- Absorber Pressure: 15-20 PSIG
  Stripper Pressure: 2-215 PSIA
- 93% of HCN and 5 to 7% of CO_2 are also removed with H_2S.

ZINC OXIDE

UNITED CATALYSTS, INC., LOUISVILLE, KENTUCKY
KATALCO, OAKBROOK, ILLINOIS

- An adsorption process, primarily used as a sulfur guard, reduces H_2S to 0.2 to 0.5 ppm level before the gas is contacted with a sulfur-sensitive (e.g., nickel-based methanation) catalyst.
- Temperature: 660 to 840°F
  Pressure: Atmospheric to over 720 PSIG
- Mercaptans and about 60% of thiophenes are also removed at about 600 to 750°F.
- More than 100 commercial plants are in operation worldwide.

STATUS SUMMARY

SELEXOL ACID GAS REMOVAL

1.0 GENERAL INFORMATION

Developer:	Allied Chemical Corporation Morristown, New Jersy
Licensor:	Norton Chemical Process Products P. O. Box 350 Akron, Ohio 44309
Type:	A physical absorption process which uses the dimethyl ether of polyethylene glycol as the solvent to remove acid gases such as CO_2, H_2S, COS and mercaptans from a variety of feed gases, including natural gas as well as synthesis gases from coal gasification and partial oxidation units.
Conditions:	The absorbers have been designed and operated between 370 and 1400 psia with acid gas concentration ranging between 5 and 65 vol%. The absorber operating temperature range is 20-100°F.
Applications:	The process has been applied to: (i) Selective removal of H_2S from raw synthesis gas generated in coal gasification or heavy oil partial oxidation processes. (ii) Selective H_2S removal or bulk CO_2 removal from sour natural gas. (iii) Removal from synthesis gas produced via reforming of natural gas and/or fuel oil. (iv) Removal of CO_2 from landfill gas to upgrade its heating value.
Process Schemes:	Many variations of the process are possible depending on the specific application. For typical coal-to-SNG applications, the most often used scheme is the selective removal of H_2S. Exact process configuration depends on the raw gas conditions, downstream processing and integration with sulfur recovery systems.
Status:	About 30 commercial and pilot installations with capacities up to 275 MMSCFD have been operating or are under construction. A list of these installations is included in Section 7.0.

2.0 PROCESS DEVELOPMENT

- In early 1960's, Allied Chemical Corporation investigated the use of dimethyl ethers of polyethylene glycols (DMPEG) for the removal of CO_2 from ammonia synthesis gas.

- In 1965, the first commercial size unit was installed at Allied Chemical's Omaha ammonia plant.

- Between 1965 and 1969, the SELEXOL process was under the following different tradenames:

  1965 - DMPEG process
  1966 - PEGASOL process
  1967 - Solvent process
  1969 - SELEXOL process

- During 1969, the SELEXOL process was first applied to the selective removal of H_2S from natural gas in the NEAG II plant in West Germany.

- In 1976, steam stripping was first used to regenerate H_2S loaded SELEXOL solvent in the Dueste II plant.

- In 1978, the SELEXOL process was installed and operated at two coal gasification pilot plants: (1) U.S. Department of Energy's BiGas Pilot plant and (2) Texaco Inc's pilot plant at Montebello, California.

- In 1982, the first commercial installation of the SELEXOL process for coal gasification application was operated at Tennessee Valley Authority's coal-to-ammonia plant. The unit is presently not operational due to lack of funding for the entire facility's operation.

- The Selexol process is also being utilized for the Cool Water coal gasification project for combined cycle application. The plant has been started up successfully according to the latest reports.

3.0 SOLVENT CHARACTERISTICS

- Physical and Chemical Properties (Source: Ref. 1 & 2)

Freeze Point	-8 to -20°F
Flash Point	304°F
Vapor Pressure @ 77°F	.0007 mmHg
Specific Heat @ 77°F	.49 BTU/lb°F
Density @ 77°F	8.57 lb/gal
Viscosity @ 77°F	5.8 cp
Thermal Conductivity, @ 77°F	.11
Surface tension @ 77°F	34.3 dynes/cm
Heat of solution @ 77°F	
CO_2	160 Btu/lb solute
H_2S	190 Btu/lb solute
CH_4	75 Btu/lb solute
Typical Molecular Weight	280

- Typical Composition (Source: Ref. 2)

Compound	Wt%
Triethylene glycol-dimethylether	12
Tetraethylene glycol - dimethylether	24
Pentaethylene glycol - dimethylether	25
Hexaethylene glycol-dimethylether	19
Heptethylene glycol - dimethylether	11
Octaethylene glycol - dimethylether	6
Nenaethylene glycol - dimethylether	3

- Bunsen Coefficients, (Source: Ref. 2)

(cm^3 gas/cm^3 solution at 68°F, 1 atm.)

H_2S	30.5
CO_2	3.4

O Relative Solubility of various gases (Source: Ref. 3)

Hydrogen		0.2
Carbon Monoxide		0.4
Methane		1.0
Ethane		6.4
Ethylene		7.3
Carbon Dioxide		15.0
Ethane		15.3
i-Butane		28.0
n-Butane		35.0
	Carbonyl Sulfide	35.0
i-Pentane		67.0
Acetylene		68.0
Ammonia		73.0
n-Pentane		83.0
	Hydrogen Sulfide	134.0
n-Hexane		165.0
	Methyl Mercaptan	340.0
n-Heptane		360.0
	Carbon disulfide	360.0
	Sulfur Dioxide	1,400.0
Benzene		3,800.0
	Thiophene	8,100.0
Water		11,000.0
Hydrogen Cyanide		18,000.0

4.0 PROCESS DESCRIPTION

Depending on the process application, many different process schemes of the SELEXOL process are used in a coal-to-SNG plant. Typically, the removal of acid gas is accomplished in two stages: selective H_2S removal followed by bulk CO_2 removal as shown in Figure 4-1.

In the first stage, H_2S is selectively absorbed from the raw synthesis gas using a lean Selexol solution. The H_2S-loaded Selexol solution is flashed to a lower pressure in a series of flash steps to remove the less soluble components from the rich solution. The flashed vapors are recycled to the absorber. The liquid is then sent to an H_2S stripper, which is used to strip out the dissolved gases consisting mainly of H_2S and CO_2. The heat input to the stripper is provided by a steam-heated reboiler. Stripper off-gases are sent to the Claus sulfur recovery unit, while the cooled lean solution is recycled to the H_2S absorber.

Synthesis gas from the H_2S absorber is further processed in the CO_2 removal section to reduce its CO_2 concentration to typically 1 vol. %, in the order to make it suitable for feeding to a methanation unit. This 1% CO_2 concentration may have to be adjusted to a level such that specifications for pipeline gas are met after methanation.

The synthesis gas leaving the H_2S absorber is cooled against CO_2 absorber overhead and ammonia refrigerant before entering the CO_2 absorber. The gas is contacted with the lean Selexol solution, giving a synthesis gas containing approximately 1 vol.% CO_2, which is sent to the methanation section. Rich solution from the CO_2 absorber is flashed to a lower pressure in a series of flash steps. Vapors from the high pressure flash are recycled and mixed with the feed gas while those from a medium pressure flash consist of high purity CO_2 gas, which can be either vented or used as lockhopper gas. The Selexol solution from the low pressure flashdrum which still contains a significant quantity of CO_2 is then sent to a CO_2 stripper to reduce the CO_2 level in the lean Selexol to less than 1 vol.%. Nitrogen or steam is used as the stripping gas. The overhead gas from the stripper is combined with gases flashed off the low pressure flashdrum and sent to a catalytic incinerator to combust the hydrocarbons present in this gas and is then vented to the atmosphere. The lean solution is recycled back to the CO_2 absorber.

A packaged refrigeration unit is provided to satisfy the requirements for chilling in the CO_2 removal stage.

An alternate process scheme (Fig.4-2) for the simultaneous removal of H_2S, COS and CO_2 is sometimes used when it involves the removal of a large amount of CO_2 and a relatively small amount of H_2S stream i.e., CO_2/H_2S ratio in the feed is high.

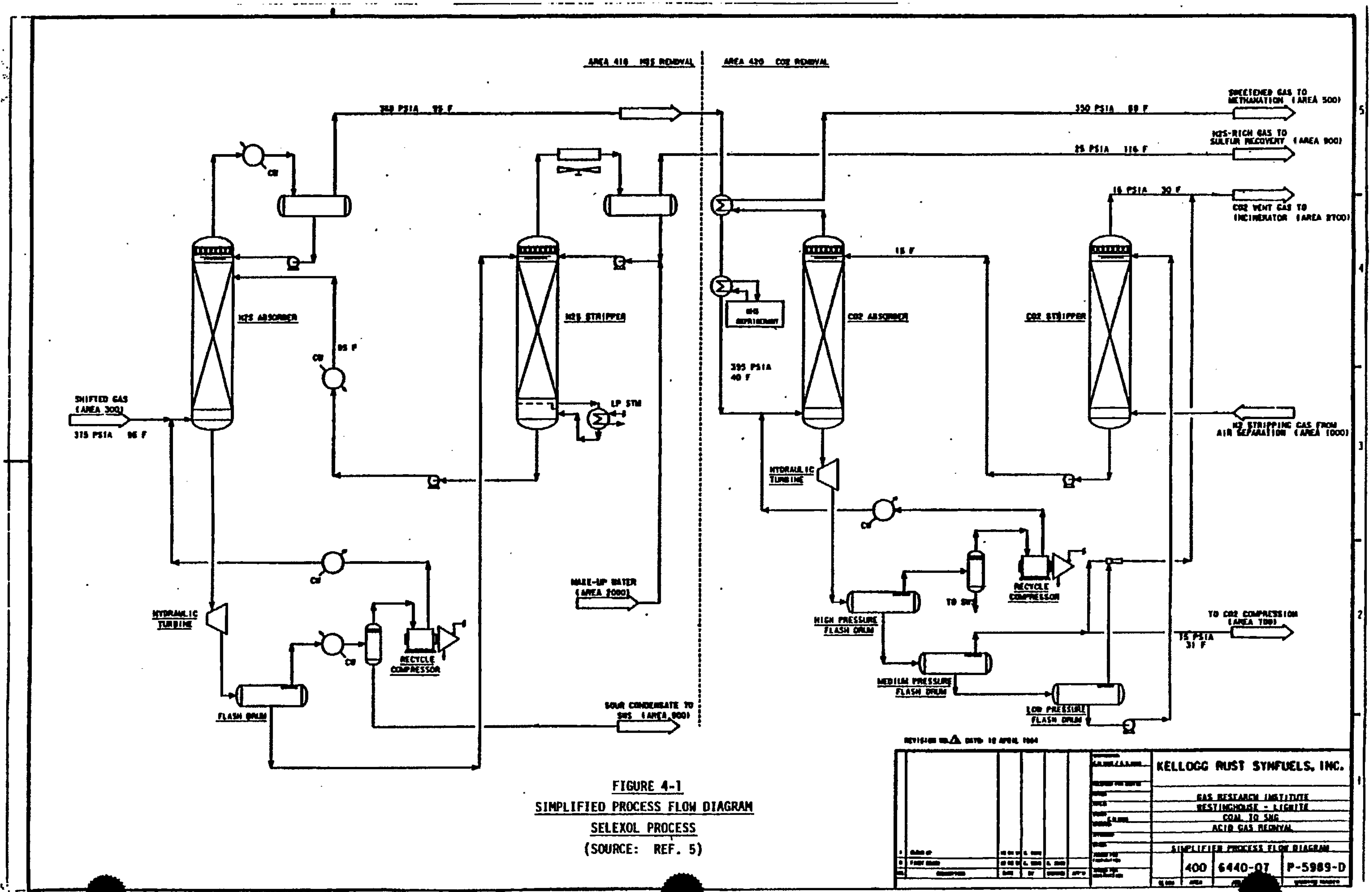

FIGURE 4-1
SIMPLIFIED PROCESS FLOW DIAGRAM
SELEXOL PROCESS
(SOURCE: REF. 5)

However non-selective, this process cannot produce a concentrated H_2S stream for Claus processing. In this split cycle scheme, all the acid gases, H_2S, COS and CO_2 are removed in a single absorber. The absorber has two sections: a lower section, fed by partially regenerated solvent, and an upper section, fed by fully regenerated solvent. COS and CO_2 are removed in the lower section with solvent which has only been flash-regenerated, and H_2S is removed in the upper section with fully regenerated solvent. The solvent flow in the upper section is much lower than in the lower section. Thus, the cost of regeneration is much less as compared to a selective scheme where the whole stream is fine-regenerated. This split-cycle scheme was employed at the TVA Coal-to-Ammonia facility.

5.0 COMMERCIAL DESIGN DATA

A summary of the design data for a selective Selexol acid gas removal system for a 250 Billion BTU/day coal-to-SNG plant is shown in Table 5-1. The feed gas is from the gasification of N. Dakota Lignite in Westinghouse gasifiers.

A set of operating data for a non-selective Selexol acid gas removal system for the TVA coal-to-ammonia facility is shown in Table 5-2. The feed gas is from the gasification of Illinois No. 6 coal in a Texaco gasifier.

Table 5-3 gives the calculation of contribution to levelized cost of gas for a selective Selexol system using data given in Table 5-1.

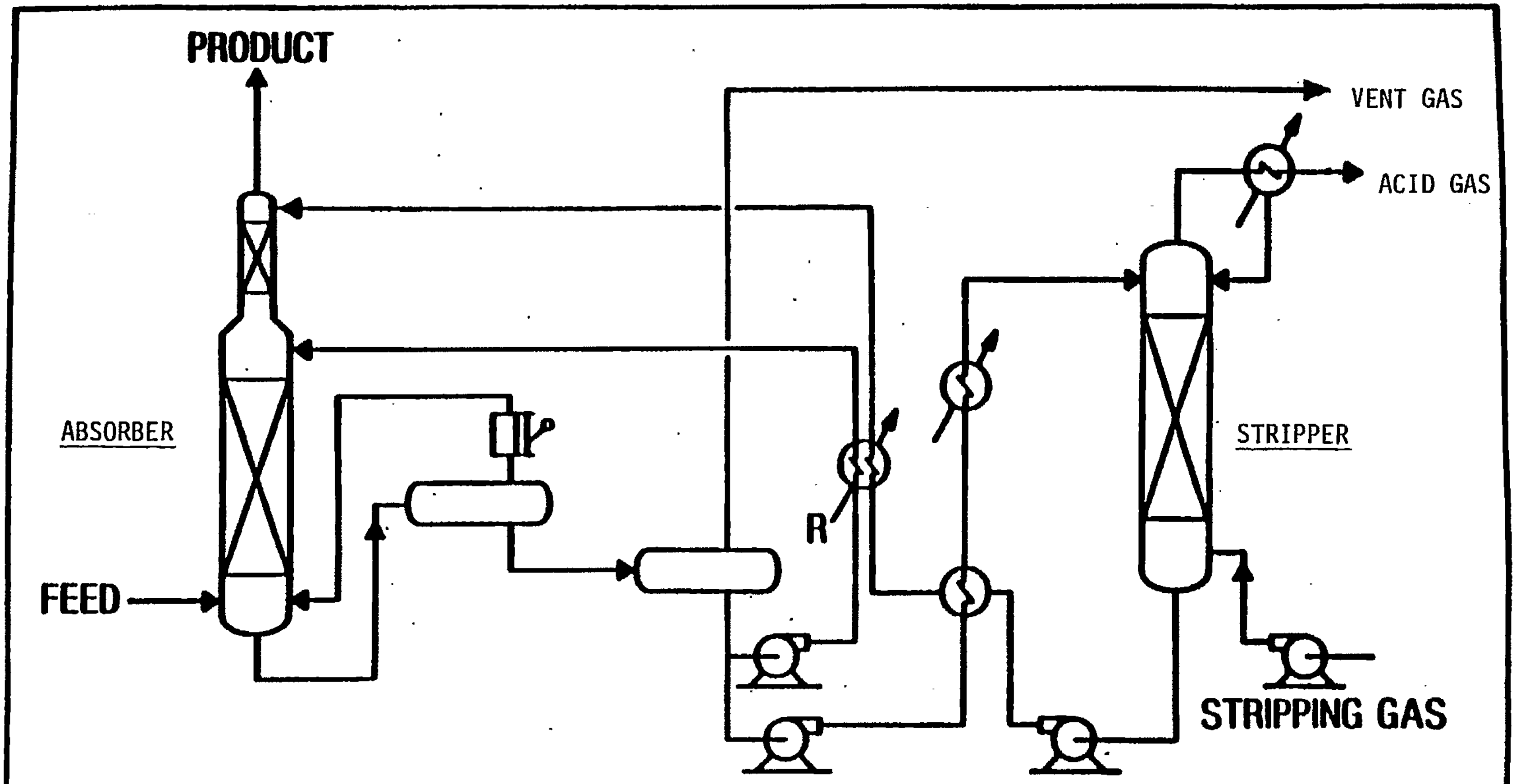

FIGURE 4-2

SIMPLIFIED ALTERNATE PROCESS FLOW DIAGRAM

(SOURCE: REF. 3)

TABLE 5-1 DESIGN DATA FOR A SELEXOL ACID GAS REMOVAL SYSTEM IN A 250 BILLION BTU/DAY COAL-TO-SNG PLANT

STREAM	FEED GAS	N2 STRIPPING GAS	MAKE-UP WATER	SWEETENED GAS	MEDIUM PRESSURE CO2	CO2 VENT	H2S-RICH GAS	SOUR CONDENSATE
	VOL%	VOL%	WT%	VOL%	VOL%	VOL%	VOL%	WT%
COMPONENT:								
CARBON MONOXIDE	13.20	0.00	0.00	20.92	0.34	0.07	0.01	0.00
HYDROGEN	42.20	0.00	0.00	67.08	0.14	0.03	0.00	0.00
CARBON DIOXIDE	37.00	0.00	0.00	1.08	98.51	80.26	69.96	0.02
METHANE	6.52	0.00	0.00	10.16	0.94	0.22	0.04	32 PPM
ARGON	0.08	0.00	0.00	0.12	0.00	0.00	0.00	0.00
NITROGEN	0.40	100.00	0.00	0.63	0.00	19.12	0.00	0.00
HYDROGEN SULFIDE	0.38	0.00	0.00	2 PPM	13 PPM	28 PPM	23.94	0.01
CARBONYL SULFIDE	6 PPM	0.00	0.00	2 PPM	7 PPM	9 PPM	37 PPM	32 PPM
WATER	0.22	0.00	100.00	0.01	0.07	0.30	6.05	99.97
TOTAL FLOW MOLES/HR	134,654	9,713	1,725	84,264	6,748	50,780	2,135	2,165
TEMPERATURE F	95	77	110	95	31	17	116	95
PRESSURE PSIA	375	18	25	368	75	20	25	55

UTILITIES CONSUMPTION:

Electricity	1,665 KW
Steam Drivers, Net	15,300 KW
Steam @ 60 psig	374,000 Lbs/Hr
Cooling Water, 15 F avg DT	38,000 Gpm
NH3 Refrigeration @ 20 F	63 MMBTU/Hr
Solvent Make-up	28 Gals/Hr

TOTAL EQUIPMENT INSTALLED COST	$ 114.8 MM (Mid-1982)
ROYALTIES	$ 1.3 MM (Mid-1982)

Source : Ref. 5

TABLE 5-2

DESIGN DATA FOR A NON-SELECTIVE SELEXOL AGR SYSTEM AT TVA COAL-TO AMMONIA PLANT

Stream	Feed gas	Acid gas to Holmes-Stretford	Sweetened Gas (Note 1)	CO_2 Product
Flowrate, Moles/hr	1,966	464	1,718	139
Pressure, psia	450.0	25.0	412.0	38.0
Temp, deg F	58.0	105.0	655.0	105.0
Composition (Vol %)				
CO	2.70	0.00	2.60	0.30
H_2	56.60	1.20	63.40	1.80
N_2/AR	0.50	11.40	20.80	0.00
CO_2	39.20	83.80	13.20	97.30
COS	8 ppm	33 ppm	< 1 ppm	13 ppm
H_2S	0.79	3.40	< 1 ppm	0.60
H_2O	0.21	0.20	0.00	0.00

Note: 1. Product gas composition from SELEXOL AGR after addition of 353 lbmoles/hr N_2 and preheated to ZnO beds feed temperature.

TABLE 5-3

CALCULATION OF CONTRIBUTION TO GAS COST
SELEXOL (SELECTIVE) AGR SYSYTEM

Coal Type	N.Dakota lignite (Westinghouse Gasifiers)
Evaluator	Kellogg Rust Synfuels,Inc.
Project Report No.	Contract No. 5082-222-0754(6440-07)
Date Published	To be published
Plant Capacity	250 Billion Btu/day SNG

CAPITAL COSTS :	$ MM (Mid-1982)
Installed Equipment	114.8
Contingency @ 15%	17.2
Direct Facility Constr Investment	132.0
Home-Office costs @ 12%	15.8
Total Facility Constr Investment	147.8
Royalties	1.3
Initial Solvent Charge	14.0
Total Plant Investment	163.1

OPERATING COSTS :			$/hr
Steam(60 psig)	374,000 #/hr	@ $ 3.85/ 1000 lb.	1439.9
Electricity	22,000 Kw	@ $0.05/Kwh	1100.0
Cooling water	38,000 Gpm	@ $0.10/ 1000 Gal	228.0
Solvent Make-up	28 Gal/hr	@ $11.81/Gal	330.7
TOTAL			3098.6

Total Operating Cost, $ MM/yr at 100 % Stream factor = 27.1 MM $/Yr

CONTRIBUTION TO GAS COSTS :	Specific Cost, $/MM Btu-Yr	Charge Rate, Year	Contribution, $/MM Btu
Capital Related	1.99	0.089	0.18
Operating	0.33	1.000	0.33
Total			0.51

6.0 ADVANTAGES/DISADVANTAGES

- Advantages

  - The SELEXOL solvent is 100% organic, biodegradable, essentially non-toxic, and non-fouling.
  - The SELEXOL solvent is chemically and thermally stable; the installation of a solvent reclaimer is not necessary.
  - The SELEXOL solvent has a very low corrosion rate on mild steel; the process operates at or near ambient conditions, hence essentially all the equipment can be of carbon steel construction
  - High H_2S selectivity with respect to CO_2 can provide a rich H_2S stream to minimize cost and increase efficiency of Claus sulfur recovery plant.
  - Simultaneous dehydration and CO_2 removal from the feed gas to produce pipeline quality gas.

- Disadvantages

  - The SELEXOL solvent has a relatively high solubility for hydrocarbons heavier than ethane.
  - In the presence of H_2S, solvent regeneration by air stripping can cause sulfur deposition. The relative solubility of H_2S in SELEXOL is approximately 4 times that of COS, so most of the COS will be removed together with CO_2. This might cause a problem if a sulfur-free CO_2 stream is required in the process. Also, the COS content of the CO_2-rich stream may not allow it to be vented directly to the atmosphere. Depending on the economics, there are two possible solutions: 1) a COS hydrolysis reactor to convert COS to H_2S upstream of the SELEXOL plant; or 2) design the absorber for COS removal, which would result in a higher SELEXOL circulation rate and a lower H_2S concentration in the acid gas stream. In certain cases, an incinerator to combust COS to SO_2 may be sufficient, since environmental regulations typically permit more SO_2 to be vented as compared to COS.

- Hydrocarbon Losses for Selective SELEXOL Process (Ref. 5 & 6)

Component	Losses, as % of Feed
CO	0.3 - 0.5
H_2	<0.1
CH_4	0.5
$C_nH_m(C_2+)$	5.6

7.0 COMMERCIAL INSTALLATIONS

There are 30 Selexol plants that are operating or are under construction for various applications including natural gas treatment, synthesis gas purification, CO_2 for Enhanced Oil ecovery, coal gasification, and landfill gas treatment.

A list of these installations is shown below in Table 7-1.

8.0 REFERENCES

1. Clare, R. T. and J. P. Valentine, "Acid Gas Removal using the SELEXOL Process"; Paper presented at the 2nd Quarterly Meeting of the Canadian Natural Gas Processors Association, June 1975, Edmonton, Alberta.

2. Woelfer, W., "Construction and Operation of a SELEXOL Natural Gas Purification Plant"; Paper presented at the 27th Canadian Chemical Engineering Conference, October 1977, Calgary, Alberta.

3. Sweeny, J. W., "Gas Treating with a Physical Solvent"; paper presented at the AIChE 1984 Winter National Meeting, March 1984, Atlanta, Georgia.

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5. Cover, A. E. et al., "Design and Economics of a Lignite-to-SNG Facility using Westinghouse Gasifier", July 1984, Gas Research Institute Contract No. 5802-222-0754.

6. Cover, A. E. et al., "Design and Economics of a Lignite-to-SNG Facility using Lurgi Gasifiers", July 1984. Gas Research Institute Contract No. 5802-222-0754.

7. Waitzman, D. A. et al., "Progress Report on Ammonia Production from Coal by Utilization of Texaco Gasification Process"; paper presented at the 1983 Fall Annual Meeting of the AIChE, November 1983, Washington, D. C.

TABLE 7-1
(SOURCE: REF. 4)

NORTON
CHEMICAL PROCESS PRODUCTS

Selexol® solvent process installations

Bulk CO_2 Removal — Synthesis Gas

Start Up	Owner & Location	Contractor	Description
1965	Allied Corporation Omaha, Nebraska	Allied Corporation Morristown, New Jersey	CO_2 from 18% to 0.5% CO_2 purity 99.5% to urea 45 MMSCFD @ 565 psia Air stripping, no steam required. Carbon steel. Original solvent charg Reforming of natural gas to NH_3
1977		Allied Corporation Morristown, New Jersey	Expansion to 69 MMSCFD
1975	Ammoniaque Synthetique et Derives S.A. (ASED) Willebroek, Belgium	Davy Powergas GmbH Cologne, W. Germany	CO_2 from 33% to <1% Operation to 50 ppm CO_2 at reduced capacity. H_2S from 200 ppm to <4 ppm HCN from 100 ppm to <10 ppm 47 MMSCFD @ 370 psia Partial oxidation of fuel oil Replaced H_2O wash
1982	Shell Oil Company Carson, California	M. W. Kellogg, Houston, Texas/ C. F. Braun, Alhambra, California	CO_2 from 19% to 200 ppm Hydrogen purification from hydrocarbon reforming Air stripping 48 MMSCFD @ 940 psia
1983	Sherritt Gordon Mines Fort Saskatchewan Alberta, Canada	M. W. Kellogg Houston, Texas	CO_2 from 18% to <0.1% New low energy ammonia plant 70% recovery of CO_2 at 98% + purity 130 MMSCFD @ 465 psia
1984	C-I-L Inc. Courtright, Ontario Canada	ICI, England/ Uhde, Dortmund, West Germany	CO_2 from 18% to <0.1% New low energy ammonia plant <25 MM Btu/Short ton Air stripping 150 MMSCFD @ 447 psia
1984	UCAM Geleen, Netherlands	Kellogg Continental Amsterdam, Netherlands	CO_2 from 18.17% to <0.10% 99% purity CO_2 to urea plant New low energy ammonia plant Air stripping 178 MMSCFD @ 468 psia

Bulk CO_2 Removal — Natural Gas

Start Up	Owner & Location	Contractor	Description
1969	Valero Transmission Co. Six Shooter, Texas Grey Ranch Plant	Fish Engineering Houston, Texas	CO_2 from 43% to <3.5% H_2S from 18 ppm to <4 ppm H_2O from saturation to <7#/MMSCF 275 MMSCFD @ 1000 psia No stripping, no refrigeration.
1974	Lone Star Gas Co. Ft. Stockton, Texas Pikes Peak Plant	Davy Powergas, Inc. Houston, Texas	CO_2 from 43% to <3.5% H_2S from 30 ppm to 4 ppm Maintains H_2O @ <7#/MMSCF 50 MMSCFD @ 1000 psia No stripping, no refrigeration.
1975		Davy Powergas, Inc. Houston, Texas	Expansion to 100 MMSCFD
1977	Northern Natural Gas Co. Ft. Stockton, Texas Mitchell Plant	Ortloff Corp. Midland, Texas	CO_2 from 28% to <3.5% H_2S from 16 ppm to <4 ppm H_2O from saturation to <7#/MMSCF 144 MMSCFD @ 900 psia No stripping, vacuum regeneration
1980	Columbia Gas Trans. Co. Kanawha, W. Virginia	Stearns Roger Inc. Denver, Colorado	CO_2 from 65% to <3% H_2S from 16 ppm to <4 ppm H_2O from saturation to <7#/MMSCF 30 MMSCFD @ 800 psia No stripping, vacuum regeneration
1979	Getty Synthetic Fuels Monterey Park, California	Ortloff Midland, Texas	CO_2 from 45% to <3% H_2O from saturation to <7#/MMSCF 8 MMSCFD @ 500 psia Landfill gas to pipeline
1980	Getty Synthetic Fuels Chicago, Illinois	Ortloff Midland, Texas	CO_2 from 38% to <3% H_2S from 8 ppm to <5 ppm H_2O from saturation to <7#/MMSCF 5 MMSCFD @ 375 psia Landfill gas
1982	Getty Synthetic Fuels Staten Island, New York	C-E Lummus Bloomfield, New Jersey	CO_2 from 43% to <3½% H_2S from 8 ppm to <5 ppm H_2O from saturation to <7#/MMSCF 15 MMSCFD @ 450 psia Landfill gas
1981	Shell Canada Resources Cranberry Field Alberta, Canada	Cord Projects Ltd. Calgary, Alberta Canada	CO_2 from 3.4% to <2% H_2S from 100 ppm to 16 ppm H_2O to <4 lbs/MMSCF 15 MMSCFD @ 1264 psia
1984	Marathon Oil U.K. Brae Field, North Sea	Matthew Hall Ltd. London, England	CO_2 from 35% to <2% H_2S from 85 ppm to <4 Vacuum regeneration 127 MMSCFD @ 700 psia

Selective Sulfur Removal — Natural Gas

Start Up	Owner & Location	Contractor	Description
1969	Northern Natural Gas Co. Ft. Stockton, Texas Oates Plant	Fish Engineering Houston, Texas	H_2S from 100 ppm to <4 ppm CO_2 from 18% to <2.5% Maintains H_2O @ <7#/MMSCF 130 MMSCFD @ 1000 psia Flash gas stripping — split flow

Selective Sulfur Removal — Natural Gas (Continued)

Start Up	Owner & Location	Contractor	Description
1970	Norddeutsche-Erdgas-Aufbereitungs-GmbH Mobil, Esso, Shell Sulingen, West Germany NEAG II Plant	Davy Powergas GmbH Cologne, W. Germany	H_2S from 4000 ppm to <2 ppm CO_2 from 7% to >5% COS from 130 ppm to 70 ppm Maintains H_2O @ <7#/MMSCF 62 MMSCFD @ 1000 psia CH_4 gas stripping Replaced potassium carbonate
1973		Mobil Oil A.G. West Germany	H_2S from 9.2% to <2 ppm $CO_2$from 9.5% to 8% COS from 130 ppm to 70 ppm RSH from 100 ppm to <1 ppm 80% H_2S to Claus plant Maintains H_2 @ <7#/MMSCF 62 MMSCFD @ 1090 psia CH_4 gas stripping Replaced Alkazid
1975		Mobil Oil A.G. West Germany	Expansion to 67 MMSCFD
1970	Wintershall AG Barnstorf, W. Germany Dueste I Plant	Davy Powergas GmbH Cologne, W. Germany	H_2S from 7% to <1000 ppm CO_2 from 9% to 5% Maintains H_2O @ <7#/MMSCF 32 MMSCFD @ 1000 psia Inert gas stripping Replaced propylene carbonate and glycol-amine
1973		Wintershall AG	Expansion to 45 MMSCFD 15 MMSCFD to <2 ppm H_2S
1976	Wintershall AG Ruetenbrock, W. Germany	Comprimo b.v. Amsterdam, Netherlands	H_2S from 60 ppm to <2 ppm CO_2 from 5% to 4% Maintains H_2O @ <7#/MMSCF 45 MMSCFD @ 1400 psia Inert gas stripping
1977	Wintershall AG Barnstorf, W. Germany Dueste II Plant	Comprimo b.v. Amsterdam, Netherlands	H_2S from 7% to <2 ppm CO_2 from 9% to 5% COS from 118 ppm to 60 ppm RSH from 100 ppm to <1 ppm 68% H_2S to Claus plant Maintains H_2O @ <7#/MMSCF 45 MMSCFD @ 1000 psia Steam stripping
1979		Comprimo b.v. Amsterdam, Netherlands	Expansion to 90 MMSCFD
1980	PanCanadian Petroleum Ltd. Alberta, Canada Morley Gas Field	MHG and Propak Calgary, Canada	H_2S from 6.7% to <8 ppm H_2O removal to <4#/MMSCF Hydrocarbon dewpoint is 15°F @ 800 psig 12 MMSCFD @ 965 psia Steam stripping 89% H_2S to Claus plant
1983	TPAO Dodan, Turkey	Williams Brothers Tulsa, Oklahoma	H_2S from 0.37% to 50 ppm 60 MMSCFD @ 700 psia
1983	Westates Italo Company Campo Reggente, Foggia, Italy	Tecnimont Milan, Italy	H_2S from 0.1% H_2S to 1.3 ppm Total Sulfur from 0.1525% to 33 ppm CO_2 from 5% to 4.45% 15.9 MMSCFD @ 1152 psia

Selexol

Selective Sulfur Removal — Synthesis Gas

Start Up	Owner & Location	Contractor	Description
1976	ERDA/AGA Homer City, Pennsylvania Bi-Gas Pilot Plant	Stearns Roger Inc. Denver, Colorado	H_2S from 7000 ppm to <4 ppm CO_2 from 31% to <1% 35% H_2S to Claus plant <20 ppm H_2S to CO_2 vent 11 MMSCFD @ 1500 psia max. Steam stripping Gasification of coal to SNG
1978	Texaco, Inc. Montebello, California Pilot Plant	Texaco, Inc. Montebello, Calif.	93.5% H_2S, COS removed Off gas combusted 1.2 MMSCFD Steam stripping Coal gasificaiton to gas turbine
1980	Tennessee Valley Auth. Muscle Shoals, Alabama	Brown & Root Houston, Texas	CO_2 from 42% to 19½% H_2S, COS from 1% to <1 ppm Sulfur off-gas to Stretford CO_2 purity to 99% to urea 18 MMSCFD @ 445 psia Nitrogen stripping Texaco coal gasification to NH_3
1983	Osaka Gas Company Osaka, Japan	Ube Industries Tokyo, Japan	Pilot plant 2.5 MMSCFD
1984	Cool Water Coal Gasification Program Daggett, California	Bechtel Houston, Texas	0.23% MSCFD, H_2S and COS to 73 ppm 19.8%, CO_2 to 15.3% 71.6 MMSCFD @ 513 psia
1983	Ministry of National Planning of Somalia Mogadishu, Somalia	Technipetrol Rome, Italy	35% CO_2 to 0.1% CO_2 0.86% H_2S to 1 ppm total sulfur 12.2 MMSCFD @ 600 psia 98.5% CO_2 stream to urea Nitrogen stripping Acid Gas to incineration
1984	International Coal Refining Company Daviess County, Kentucky	Ralph M. Parsons Pasadena, California	H_2S from 1.3% to 1 ppm COS from 60 ppm to 3 ppm CO_2 from 41% to 0.2% 137 MMSCFD @ 720 psia 26% H_2S to Claus plant
1984	Electric Power Development Company, Japan	Kawasaki Heavy Industries Kobe, Japan	Pilot Plant 0.6 MMSCFD

Pending Installation and Major Studies

Natural gas treatment on an offshore platform.

Coal gasification to low Btu industrial gas and combined cycle power generation.

Coal liquefaction programs.

CO_2 for enhanced oil recovery.

Grass roots low energy ammonia plants.

Coal or peat to methanol projects.

NORTON

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STATUS SUMMARY

RECTISOL ACID GAS REMOVAL

1.0 GENERAL INFORMATION

Developers:	Lurgi Gesellschaft fur Waermetechnik mhH, Frankfurt am Main, Federal Republic of Germany and Linde Aktiengesellschaft, Munchen, Federal Republic of Germany.
Licensors:	Lotepro Corporation 1140 Avenue of the Americas New York NY 10036
Type:	A physical absorption process which uses methanol as the solvent to remove acid gases (CO_2, H_2S and COS)and other impurities (NH_3, HCN, gumformers and higher hydrocarbons) from synthesis gases produced via coal gasification or heavy oil partial oxidation.
Conditions:	Temperatures are maintained in the range of -100 to 0°F in the absorbers, and up to approximately 150°F on the regenerator with steam reboiler. The process has been designed and operated between 315 to 2500 psia with acid gas concentration ranging between 5 and 45 vol%.
Applications:	Rectisol has been applied to: (i) removal of CO_2, H_2S, COS, NH_3, HCN, gumformers higher hydrocarbons and other impurities from gases produced via coal gasification. (ii) Removal of CO_2, H_2S, COS, HCN from synthesis gas produced by partial oxidation of hydrocarbons.
Process Schemes:	Several variations of process are possible depending on the specific application for a typical coal-to-SNG application the most often used schemes are (1) selective removal of H_2S and (2) Non-selective or bulk removal of acid gases. Selection of schemes depends on several factors including coal type, integration with sulfur recovery system and downstream processing.

Status: More than 70 units are in operation or under contruction for the production of synthesis gas, H_2 or towngas mainly based in either partial oxidation of oil or coal gasification. The Rectisol process has been used at the Sasol plants and more recently in the Great Plains Coal Gasification Associates' coal-to-SNG plant in North Dakota, and the coal-to-chemicals plant by Tennessee Eastman in Tennessee.

2.0 PROCESS DEVELOPMENT

Prior to 1950, Lurgi and Linde had collected laboratory data on the solubilities of various gases in methanol. In 1950, a 200 Nmhr pilot plant was erected by Lurgi and Linde at Ruhrchemie AG's Oberhausen-Holten plant site.

The Rectisol Process was first commericialized in 1957 at Sasol I. Three scrubbing units, each with 60,000 Nm^3/hr raw gas capacity, were installed.

3.0 SOLVENT CHARACTERISTICS

- Physical and Chemical Properties

Freeze Point	-143.0 °F	
Flash Point	1.2°F	
Molecular Weight	32.0	
Vapor Pressure @ -60oF	0.5	mm Hg
Specific Heat, liquid * @ -60°F	0.57	Btu/lb°F
Density @ -60°F	7.14	lb/gal
Viscosity* @ -60°F	2.50	CP
Thermoconductivity @ -60°F	0.13	Btu/hr-ft°F
Surface tension* @ -60°F	28.50	dynes/cm

- Solubilities of gases in methanol at -60°F

	LB mols (gas in solution)/ Tons (methanol)/atm. (partial pressure of gas)	Relative Solubilities
H_2	0.006	0.071
N_2	0.006	0.071
CO	0.032	0.381
CH_4	0.084	1.000
CO_2	4.229	5.035
COS	15.000	178.571
H_2S	29.900	355.952

*Approximate Readings (Reference 1)

4.0 PROCESS DESCRIPTION

Depending on the process application, many different process schemes of the Rectisol Process can be used. In a typical coal to SNG plant, the H_2S, CO_2 and COS can either be selectively removed or simultaneously removed. The feed gas composition, the downstream processing, and the sulfur recovery system should all be considered in the process scheme selection.

NON-SELECTIVE RECTISOL PROCESS

Figure 4-1 describes a non-selective H_2S, COS, and CO_2 removal scheme for processing a feed gas stream originating from Lurgi gasifiers.

- Feed Gas Cooling and Prewash - The feed gas is cooled by the purified synthesis gas, flash gases and refrigerant. The condensed gas liquor is collected in condensate separators. The cooled feed gas enters the prewash tower where it is washed with a small quantity of methanol to remove the naphtha, HCN, and water.

- H_2S, COS and CO_2 Removal - The main gas stream leaving the prewash tower is scrubbed in the main wash section of the H_2S absorber where most of the H_2S, COS, and CO_2 are removed using flash regenerated methanol. An intermediate chilling loop in the middle section of the absorber is used to remove the heat of absorption. From the main section, the gas flows to the fine wash section in the top of the H_2S absorber where it is scrubbed with fully regenerated methanol to remove the remaining sulfur compounds and CO_2 to the specified level. The sweet gas then exits the Rectisol unit by heat exchange with the feed gas.

- Flash Regeneration - The rich methanol leaving the bottom of the main section of the absorber is regenerated in the flash regenerator. The laden methanol is flashed in several successive stages down to a slight vacuum, where dissolved gases are released. Gas from the first flash stage is recycled to the gasification section and used as lockhopper gas so that the bulk of the co-absorbed CH_4, CO and H_2 are not lost. The other flash gases are sent to a sulfur recovery unit that can handle acid gases with a low sulfur content. A portion of the bottoms of the flash regenerator is used as the semi-lean solution used in the middle section of the absorber, while the remainder is further regenerated by steam.

- Hot Regeneration - Before the flash regenerated methanol enters the hot regenerator, it is heated up by the hot regenerated methanol. In the hot regenerator, the dissolved gases are stripped out by vapors generated by a steam heated reboiler. The fully regenerated methanol is used in the top section (fine wash section) of the

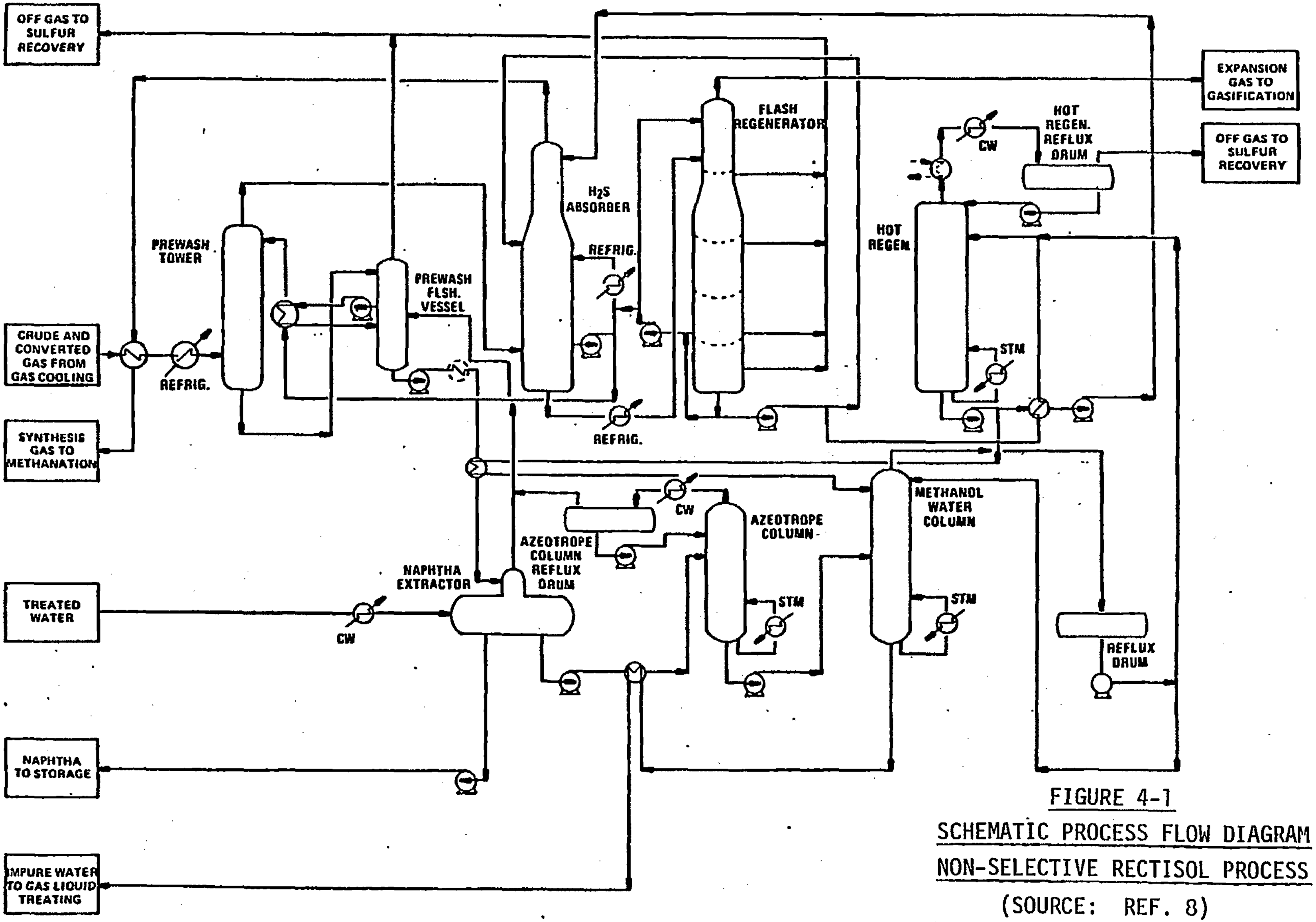

FIGURE 4-1

SCHEMATIC PROCESS FLOW DIAGRAM

NON-SELECTIVE RECTISOL PROCESS

(SOURCE: REF. 8)

absorber. The off gas leaving the hot regenerator is sent to a sulfur recovery unit.

- Naphtha and Methanol Recovery - The liquid from the bottom of the prewash tower consisting of methanol, naphtha and water is degassed in a prewash flash vessel. The gases from the flash vessel are combined with other flash gas streams going to a sulfur recovery unit, while the liquid is passed to the naphtha extractor. By addition of water, the naphtha and methanol layers are separated and crude naphtha is withdrawn and sent to storage. The methanol-water layer, which is not entirely free from naphtha, is warmed up and fed to the azeotrope column. The azeotropic mixture of naphtha and methanol is discharged from the top of the column and ultimately returned to the naphtha extractor. The water-methanol mixture is withdrawn from the bottom of the azeotrope column and fed to the methanol water column, where the water is separated from methanol by distillation. Water is drawn off from the bottom of the column, cooled and sent to waste water treatment. The regenerated methanol is then recycled back to hot regeneration column.

SELECTIVE RECTISOL PROCESS

Besides the non-selective removal of H_2S and CO_2, a process scheme for selective removal of H_2S and CO_2 is possible. Figure 4-2 shows a selective one-stage wash scheme to treat synthesis gas from a partial oxidtion unit after it has been quenched and shifted.

- Feed Gas Precooling-Feed gas is cooled by heat interchange with the product and waste CO_2 streams in exchanger E1. Icing is prevented by injecting methanol into the gas stream prior to cooling.

- Feed Gas Purification - After separation of the condensed methanol-water mixture, the feed gas is scrubbed by methanol in the wash column, T1. H_2S and COS are scrubbed by CO_2-loaded methanol in the bottom section of the absorber. The removal of CO_2 is accomplished by scrubbing with hot regenerated methanol in the top section of the absorber. The heat of solution due to CO_2 absorption is removed by external refrigeration. Purified gas leaves the top of the absorber.

- Cold Regeneration - A portion of the CO_2-loaded methanol from the absorber (T1) is flashed in successive stages (D2). The flash gases are compressed and recycled to the absorber T1, so as to recover the co-absorbed syngas. The cold regenerated methanol is fed to the top of H_2S enrichment column, T2.

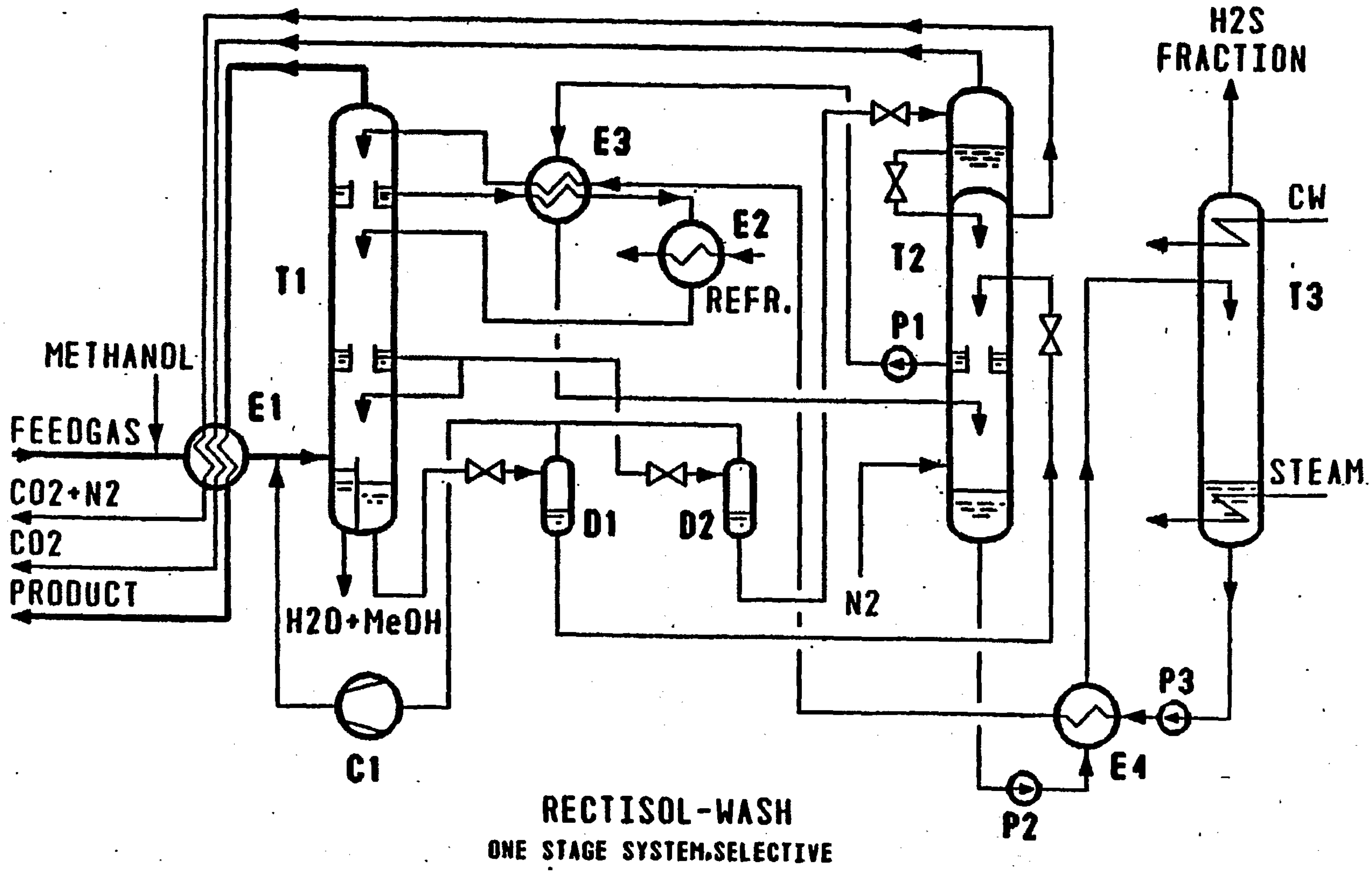

FIGURE 4-2 ONE-STAGE SELECTIVE RECTISOL WASH SYSTEM (REF. 5)

- H_2S Enrichment - The H_2S-rich methanol leaving the bottom of the absorber is flashed in a tank (D1), to an intermediate pressure. The flash gases are compressed and recycled, while the methanol solution from the flash vessel is fed into the middle of the H_2S enrichment column. In order to achieve the required concentration of H_2S in the acid gas stream, N_2 is used to strip out CO_2 from the loaded methanol in the lower section of the column. The H_2S content of the vent stream is controlled by contacting it with sulfur-free, CO_2-loaded methanol from the cold regeneration step. The sulfur-free tail gas leaves the top of the column, T_2.

- Warm Regeneration - The H_2S-rich methanol from the bottom of the H_2S concentration column is warmed up against warm regenerated methanol in exchanger E4 and pumped (P2) into the regeneration column, T3. Complete stripping of H_2S and CO_2 is affected by a steam reboiler. The H_2S-rich stream leaving the top of the regeneration column is cooled and the condensed methanol is refluxed to the top of the regeneration column. The H_2S-rich stream is sent to a sulfur recovery unit. The regenerated methanol leaving the bottom of the column is cooled in E4, before being sent to the absorber, T1.

5.0 COMMERCIAL DESIGN DATA

A summary of the design data for a "Selective" Rectisol acid gas removal system for a 250 Billion BTU/day coal-to-SNG plant is shown in Table 5-1. The feed gas is from the gasification of Kentucky #9 coal in Exxon Catalytic Gasifiers.

Table 5-2 presents the design data for a "Non-Selective" Rectisol acid gas removal system for a 125 Billion BTU/day coal-to-medium-BTU gas plant. The feed gas is from the gasification of Texas Lignite in Lurgi gasifiers.

Tables 5-3 and 5-4 give the calculation of contributions to gas cost for the two design cases presented in Tables 5-1 and 5-2.

TABLE 5-1

DESIGN DATA FOR A SELECTIVE RECTISOL AGR SYSTEM IN A 250 BILLION BTU/DAY COAL-TO-SNG PLANT

Stream *	Feed gas	Acid gas to Sulfur Recovery	Tail gas	CO_2 Vent	Sweetened Product Gas
Flowrate,					
Moles/hr	110,358	2,338	3,923	16,542	87,138
#/hr	2,032,572	83,925	163,817	727,572	1,039,758
Pressure, psia	450.0	26.2	21.8	17.4	432.0
Temp,deg F	120.0	95.0	89.3	89.3	5.0
Composition (Vol %)					
CO	9.52	0.05	1.37	0.02	11.99
H_2	34.14	0.05	3.74	0.01	43.06
CH_4	32.31	0.21	1.63	0.06	40.83
C_2H_6	45 ppm	0.01	0.02	0.01	23 ppm
N_2	3.25	0.22	0.02	0.00	4.11
CO_2	19.60	61.70	93.21	99.89	0.01
COS	18 ppm	0.08	0.00	< 1 ppm	<0.3 ppm
H_2S	0.80	37.66	0.00	0.01	<0.4 ppm
H_2O	0.38	0.00	0.00	0.00	6 ppm

UTILITY REQUIREMENTS

Electric Power	59,000 Kw
Steam	
65 psig	260,000 #/hr
150 psig	38,000 #/hr
Cooling Water	24,800 Gpm
Make-up Solvent	23,760 #/day

CAPITAL COSTS

Equipment (Installed)	$ 77.1 MM (March,1979)
Royalties	$ 3.7 MM (March,1979)

* Refer to Figure 4-2

TABLE 5-2

DESIGN DATA FOR A NON- SELECTIVE RECTISOL AGR SYSTEM IN A 125 BILLION BTU/DAY COAL-TO-MED. BTU GAS PLANT

Stream *	Crude gas from gas cooling	Off gas to Sulfur Rec	Product Gas to Methanation	Naphtha to Storage	Net Water Purge
Gas Flowrate, #/hr	1,046,100	662,349	381,317		
Gas Composition (Vol %)					
CO	15.35	0.11	22.19		
H2	40.76	0.08	59.00		
CH4	11.16	0.82	15.81		
CnHm	1.17	2.48	0.58		
N2	0.29	0.00	0.42		
CO2	30.23	94.02	2.00		
COS	0.02	0.05	0.00		
H2S	0.76	2.44	<10 ppm		
H2O	0.26	0.00	0.00		
Liquids, #/hr					
H2O					2,334
HCl/HCN	100				100
Naphtha	8,469			8,469	
Total Flowrate, #/hr	1,054,669	662,349	381,317	8,469	2,434

UTILITY REQUIREMENTS

Electric Power	6,900 Kw
Refrigeration	7,910 Kw
Steam - 100 psig	59,000 #/hr
Boiler Feed water	27,000 #/hr
Cooling Water	3,960 Gpm
Make-up Solvent	13,200 #/day
Caustic	1,776 #/d

CAPITAL COSTS

Equipment (Installed)	$ 56.2 MM (4th Quarter,1981)
Royalties	$ 4.9 MM (4th Quarter,1981)

* Refer to Figure 4-1

7

TABLE 5-3

CALCULATION OF CONTRIBUTION TO GAS COST
SELECTIVE RECTISOL CASE

Coal Type	Kentucky # 9 (Exxon Gasifiers)
Evaluator	Air Products & Chemicals,Inc.
Project Report No.	FE-3044-T12
Date Published	February,1980
Plant Capacity	250 Billion Btu/day SNG

CAPITAL COSTS :	$ MM (Mid-1982)
Installed Equipment	113.4
Contingency @ 15%	17.0
Direct Facility Constr Investment	130.4
Home-Office costs @ 12%	15.6
Total Facility Constr Investment	146.0
Royalties	5.4
Initial Solvent Charge	0.5
Total Plant Investment	151.9

OPERATING COSTS :			$/hr
Steam(65 psig)	260,000 #/hr	@ $ 3.85/ 1000 lb.	1001.0
Steam(150 psig)	38,000 #/hr	@ $4.00/ 1000 lb.	152.0
Electricity	59,000 Kw	@ $0.05/Kwh	2950.0
Cooling water	24,800 Gpm	@ $0.10/ 1000 Gal	148.8
Solvent Make-up	150 Gal/hr	@ $0.55/Gal	82.4
TOTAL			4334.2

Total Operating Cost, $ MM/yr at 100 % Stream factor = 37.9 MM $/Yr

CONTRIBUTION TO GAS COSTS :	Specific Cost, $/MM Btu-Yr	Charge Rate, Year	Contribution, $/MM Btu
Capital Related	1.85	0.089	0.16
Operating	0.46	1.000	0.46
Total			0.62

TABLE 5-4

CALCULATION OF CONTRIBUTION TO GAS COST
NON-SELECTIVE RECTISOL CASE

Coal Type	Texas Lignite (Lurgi Gasifiers)
Evaluator	Fluor
Project Report No.	DOE/RA/50381-1156
Date Published	March,1982
Plant Capacity	125 Billion Btu/day Med Btu gas

CAPITAL COSTS :	$ MM (Mid-1982)
Installed Equipment	57.6
Contingency @ 15%	8.6
Direct Facility Constr Investment	66.2
Home-Office costs @ 12%	7.9
Total Facility Constr Investment	74.1
Royalties	5.0
Initial Solvent Charge	0.1
Total Plant Investment	79.2

OPERATING COSTS :			$/hr
Steam(65 psig)	59,000 #/hr	@ $ 3.85/ 1000 lb.	227.2
Boiler Feed Water	27,000 #/hr	@ $1.50/ 1000 lb.	40.5
Electricity	14,810 Kw	@ $0.05/Kwh	740.5
Cooling water	3,960 Gpm	@ $0.10/ 1000 Gal	23.8
Solvent Make-up	150 Gal/hr	@ $0.55/Gal	42.6
Caustic	74 #/hr	@ $190/ton	7.0
TOTAL			1081.6

Total Operating Cost, $ MM/yr at 100 % Stream factor = 9.47 MM $/Yr

CONTRIBUTION TO GAS COSTS :	Specific Cost, $/MM Btu-Yr	Charge Rate, Year	Contribution, $/MM Btu
Capital Related	1.93	0.089	0.17
Operating	0.23	1.000	0.23
Total			0.40

6.0 ADVANTAGES & DISADVANTAGES

- Advantages

  - Methanol is inexpensive, generally non-corrosive and readily available.

  - Methanol is not degraded by trace components such as HCN, NH_3, CS_2, COS or other hydrocarbon impurities produced from coal gasification.

  - The process is attractive for applications requiring high purity product and for feed gases having a high acid gas partial pressure. A 0.1 ppmv total sulfur (including H_2S, COS, CS_2, mercaptans, etc.) specification in the treated gas stream can be achieved.

  - A COS hydrolysis step is not required since COS can be removed economically with the H_2S stream.

  - Approximately 50 to 70% of the CO_2 content in the feed gas can be recovered as a CO_2 product steam of 99^+% purity.

  - The "Selective" Rectisol process can produce a high H_2S concentration acid gas stream, which can be sent directly to a Claus sulfur recovery unit.

  - The product gas is dehydrated simultaneously with CO_2 removal from the feed gas.

  - The Rectisol process also serves as a precooling step if subsequent low temperature processing of the purified gas (such as liquid N_2 wash, H_2/CO separation, liquid CH_4 wash, etc.) is required.

- Disadvantages

  - The process flow scheme is very complex

  - Extensive use of stainless steel and 3.5% nickel steel due to low operating temperatures results in higher capital costs.

  - Solvent vaporization losses are relatively high due to the appreciable vapor pressure of methanol even at low temperatures. Hence, depending on the plant location and its size, a small methanol plant on site may be necessary.

- Nitrogen, which is used as a stripping gas, has to be very pure(O_2 = 3-10 ppm) due to the following reasons:
  - (i) Potential safety hazards for the methanol-O_2 mixture.
  - (ii) Elemental sulfur precipitation.
  - (iii) Safety problems associated with the presence of O_2 in the cryogenic units.

o Hydrocarbon Losses for "Non-Selective" Rectisol (Ref. 4, 6, 7).

Component	Losses, as % of Feed
CO	0.3-0.5
H_2	0.1
CH_4	0.5-3
$C_nH_m(C_3{}^+)$	55-65

7.0 COMMERCIAL INSTALLATIONS

About 70 units have been built and operated since the 1950's by both Lurgi and Linde. The Rectisol plants built by Lurgi include those at SASOL I, SASOL II, SASOL III. A list of Rectisol units built by Linde/Lotepro is shown in Table 7-1 and the units built by Lurgi are shown in Table 7-2.

8.0 REFERENCES

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4. Schreiner, Max; "Research Guidelines Studies to Assess Gasoline from coal by Methanol-to-Gasoline and SASOL-Type Fischer-Tropsch Technologies." Final Report, August 1978, FE-2447-13.

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6. Cassano, A.A. et al. "Cryogenic Methane Separation/Catalytic Hydrogasification Process Analysis", Final Report, February 1980, FE-3044-T12.

7. Transco Medium BTU Coal Gasification Project: Feasibility Study - Final Report, March 1982, Volumes I, II and III. DOE/RA/50 381-1156.8. "Great Plains Gasification Project", Final Environmental Impact Statement, Volume 1, DOE EIS-0072F, August 1980.

TABLE 7-1 (Reference 3)

Note: Three units were built by Messer Griesheim GmbH, whose know-how was acquired and transferred to Linde AG in 1971.

CLIENT	FEEDGAS	PRESSURE	COMPONENTS REMOVED (VOL.%)	CONTRACT AWARDED IN
Typpi Oy Oulu, Finland	16 MMSCFD hydrogen rich gas from partial oxidation of heavy fuel oil	356psia	CO_2,33.4% - 50ppm H_2S,0.4% - 1 ppm CO_2-Prod. 2.1 MMSCFD 99%	1960
Dansk-Norsk Kvaelstoff-fabrik, Copenhagen, DK	13.8 MMSCFD hydrogen rich gas from partial oxidation of heavy fuel oil	356 psia	CO_2 33.4% - 50 ppm H_2S 0.4% - 1 ppm CO_2 -Prod. 1.45 MMSCFD 99%.	1961
Texaco Inc. Los Angeles, USA.	79.6 MMSCFD hydrogen rich gas from partial oxidation.	481 psia	CO_2 33.3% - 10 ppm H_2S 0.49% - 1 ppm Tailgas: max. 5 ppm H_2S no CO_2 required. H_2S-Prod. 10% design result 25.6% H_2S	1966
H. Koppers GmbH, Essen, for Kutahya, Turkey.	46.9 MMSCFD converted gas from coal gasification.	351 psia	CO_2: 37.6% - 60 ppm H_2S. Traces to be removed CO_2-Prod. 12.9% MMSCFD 99%	1966

TABLE 7-1 (Continued)

CLIENT	FEEDGAS	PRESSURE	COMPONENTS REMOVED (VOL.%)	CONTRACT AWARDED IN
American Air Liquide Inc. for Monsanto Texas City, USA.	53 MMSCFD hydrogen rich gas from steam reformer	351 psia	CO_2 10.2% - 20 ppm	1969
Long Island Lighting Comp. New York, USA.	4.3 MMSCFD natural gas.	602 psia	Odorants, CO_2, C_2+ CO_2 0.92 - 5 ppm	1969
Pilot - and Demonstration Plant for Texaco Inc. Monte Bello CAL.	1 MMSCFD hydrogen rich gas from partial oxidation of heavy hydrocarbons.	1140-2560 psia	CO_2 33% - 10 ppm H_2S 0.7% - 1 ppm H_2S in tailgas max. 5 ppm	1966
Borden Chemical Comp. New York, USA.	16.8 MMSCFD	313 psia	CO_2	1966
Rohm and Haas Comp. Philadelphia, USA.	12.6 MMSCFD	356 psia	CO_2, C_2H_2	1966
Brooklyn Unio Gas Brooklyn New York, USA	11.9 MMSCFD	356 psia	CO_2	1966
Rohm GmbH, Darmstadt, West Germany-	7.9 MMSCFD natural gas.	498 psia	S-Components, C_2+ CO_2 5% - 50 ppm H_2S o.01% - 15 ppm	1970

TABLE 7-1 (Continued)

CLIENT	FEEDGAS	PRESSURE	COMPONENTS REMOVED (VOL.%)	CONTRACT AWARDED IN
H.Koppers GmbH, Essen, for Modderfontein, South Africa.	146 MMSCFD hydrogen rich gas from coal gasification.	427/711 psia	CO_2 42.6% - 20 ppm H_2S 0.95% - 1 ppm H_2S-Prod. 80% H_2S/COS CO_2-tailgas: max. 10 ppm H_2S CO_2-Prod. 10.8 MMSCFD 99.9%	1972
Krupp Koppers GmbH. Essen, for Kafue, Zambia.	32.9 MMSCFD hydrogen rich gas from coal gasification.	398/341 psia	CO_2 42.3% - 20 ppm H_2S 1.0% - 20 ppm H_2S-Prod. 97% H_2S/COS CO_2-tailgas: max. 150 ppm H_2S	1975
Celanese Chemical Co. Houston, Texas, USA	Over 35 MMSCFD hydrogen rich gas.	441 psia		1975
Syngas Co. Houston, Texas, USA	from partial oxidation of heavy residual oil before and after CO-shift (2 stages)	853 psia	CO_2, H_2S, COS, HCN - 0.1 ppm H_2S + COS	1976
Gujarat State Fertiliser Co., Baroda, India.	168 MMSCFD Converted gas from partial oxidation of heavy fuel oil (1 stage)	1067 psia	35.9% CO_2 - 20 ppm 0.75% H_2S - 1 ppm CO_2-Prod. 26.8 MMSCFD - 98.5% CO_2	1976
UBE Industries Ltd., Tokyo.	21.1 MMSCFD from partial oxidation of heavy fuel oil.	612 psia	6.7% CO_2 - 20 ppm 1.46% (H_2S + COS) - 0.1 ppm H2S-fraction: 50% (H_2S + COS)	1977

TABLE 7.1 (Continued)

CLIENT	FEEDGAS	PRESSURE	COMPONENTS REMOVED (MOL %)
Nagarjuna Fertilizers and Chemicals Ltd., Kakinada, India	116.1 MM SCFD Shifted Gas from partial oxidation of heavy fuel oil	682 psia	CO_2: 35.4% - 20 ppm H_2S + COS 0.74% - 1 ppm
China National Technical Import Corp., Beijing, PR China	3 x 121.7 MMSCFD Shifted gas from partial oxidation of vacuum residual oil	1138 psia	CO_2: 34.10% - 10 ppm H_2S: 0.24 - .1 ppm
Chinese Petroleum Corp., Taipei, Taiwan	119.3 MMSCFD Shifted and unshifted gas from partial oxidation of heavy residual oil	870 psia	CO_2: 36% - 10 ppm H_2S + COS: 1.5% - 0.1 ppm
Bechtel Petroleum for Tennessee Eastman Corp., Kingsport, Tennessee U.S.A.	44.7 MMSCFD unshifted and 25.6 MMSCFD shifted gas from Texaco coal gasification	860 psia	CO_2: 38% - 20 ppm H_2S + COS: 1.28% - 0.1 ppm
Texaco Corporation Convent, Louisiana, U.S.A,	108 MMSCFD	690 psia	CO_2 + H_2S
C. E. Lummus for Great Plains Associates North Dakota, U.S.A.		450 psia	CO_2 + H_2S

Gas- and Synthesis Technology

TABLE 7-2

2. Gas Purification by Physical Absorption

2.1 Rectisol® Process

Plant Pressure: 22 to 55 bar

Built up to 1964: a total of 9 units with a maximum plant capacity of 6.0 MM Nm^3/day

Date of Order	Location	No. of Units	Feedstock	Components to be Removed	Product for	Capacity m^3/day	Process/ Licensor
	CSSR, Germany, South Africa, USSR		Raw Gas ex Coal-, Oilgasification	CO_2, H_2S, org. S, Hydrocarbons	Town Gas, NH_3 Synthesis, Fischer-Tropsch Synthesis	15,700,000	Lurgi
1964	Republic of South Africa	1	Raw Gas ex Coal-gasification	CO_2, H_2S, org. S, Hydrocarbons	Town Gas	2,800,000	Lurgi
1964	USA	1	Raw Gas	CO_2	H_2 and CO Recovery	450,000	Lurgi
1964	USA	1	Raw Gas	CO_2, C_2H_2	H_2 and CO Recovery	340,000	Lurgi
1965	CSSR	1	Raw Gas ex Coal-gasification	CO_2, H_2S, org. S, HCN, Hydrocarbons	Town Gas	2,400,000	Lurgi
1965	Yugoslavia	1	Raw Gas ex Coal-gasification	CO_2, H_2S, org. S, HCN, Hydrocarbons	NH_3, Urea Synthesis Town Gas	2,200,000	Lurgi
1966	USA	1	Natural Gas	CO_2	Natural Gas Peak Shaving	320,000	Lurgi
1967	CSSR	2	Raw Gas ex Coal-gasification	CO_2, H_2S, HCN, Hydrocarbons	Town Gas	5,800,000	Lurgi
1967	Yugoslavia	2	Raw Gas ex Coal-gasification	CO_2, H_2S, org. S, HCN, Hydrocarbons	NH_3, Urea Synthesis Town Gas	2,500,000	Lurgi
1967	Germany	1 1	Raw Gas ex Oil-gasification	H_2S, COS, CO_2	Methanol Synthesis	1,300,000 960,000	Lurgi
1971	Germany	1 1	Raw Gas ex Oil-gasification	H_2S, COS, HCN, CO_2	NH_3, Methanol Synthesis	4,000,000 4,700,000	Lurgi
1971	Great Britain	1	Raw Gas ex Coal-gasification	CO_2, H_2S, HCN, org. S. Hydrocarbons	Substitute Natural Gas	280,000	Lurgi
1971	USA	2	Raw Gas ex Coal-gasification	CO_2, H_2S, org. S, Hydrocarbons	Substitute Natural Gas	14,900,000	Lurgi
1975	India	1 1	Raw Gas ex Oil-gasification	H_2S, COS, HCN, CO_2	NH_3, Urea Synthesis	2,100,000 2,900,000	Lurgi

Source: Lurgi literature

Gas- and Synthesis Technology

2. Gas Purification by Physical Absorption

2.1 Rectisol® Process (cont'd)

Date of Order	Location	No. of Units	Feedstock	Components to be Removed	Product for	Capacity m³/day	Process/ Licensor
1975	Republic of South Africa	1	Raw Gas ex Coal-gasification	CO_2, H_2S, HCN, org. S Hydrocarbons	Fischer-Tropsch Synthesis Town Gas	5,300,000	Lurgi
1975	Germany	1	Raw Gas ex Oil-gasification	H_2S, COS, HCN,	Methanol Synthesis	1,700,000	Lurgi
1976	India	1 1	Raw Gas ex Coal-gasification	H_2S, COS, HCN CO_2	NH_3, Urea Synthesis	2,400,000 3,600,000	Lurgi
1976	India	1 1	Raw Gas ex Oil-gasification	H_2S, COS, HCN, CO_2	NH_3-Synthesis	2,100,000 2,900,000	Lurgi
1976	India	1 1	Raw Gas ex Oil-gasification	H_2S, COS, HCN, CO_2	NH_3-, Urea Synthesis	2,400,000 3,600,000	Lurgi
1976	India	1 1	Raw Gas ex Oil-gasification	H_2S, COS, HCN, CO_2	NH_3-Synthesis	2,100,000 2,900,000	Lurgi
1976	India	1 1	Raw Gas ex Oil-gasification	H_2S, COS, HCN, CO_2	NH_3-Synthesis	2,100,000 2,900,000	Lurgi
1976	Germany	1 1	Raw Gas ex Oil-gasification	H_2S, COS, HCN, CO_2	NH_3-Synthesis	3,800,000 5,450,000	Lurgi
1977	India	1 1	Raw Gas ex Oil-gasification	H_2S, COS, HCN, CO_2	NH_3-Synthesis	1,800,000 2,200,000	Lurgi
1977	Republic of South Africa	4	Raw Gas ex Coal-gasification	CO_2, H_2S, HCN, org. S Hydrocarbons	Fischer-Tropsch Synthesis	39,600,000	Lurgi
1977	Japan	1 1	Raw Gas ex Oil-gasification	H_2S, COS, HCN, CO_2	Methanol Synthesis	1,100,000 800,000	Lurgi
1978	Brazil	1	Raw Gas ex Oil-gasification	H_2S, COS, HCN, CO_2	NH_3-, Urea Synthesis	2,700,000 3,900,000	Lurgi
1978	China	1	Raw Gas ex Coal-gasification	CO_2, H_2S, COS, org. S Hydrocarbons	NH_3-, Urea Synthesis	2,880,000	Lurgi
1979	Portugal	1 1	Raw Gas ex Oil-gasification	H_2S, COS, HCN, CO_2	NH_3-, Urea Synthesis	2,400,000 3,200,000	Lurgi

2. Gas Purification by Physical Absorption

2.1 Rectisol® Process (cont'd)

Date of Order	Location	No. of Units	Feedstock	Components to be Removed	Product for	Capacity m³/day	Process/ Licensor
1979	Republic of South Africa	4	Raw Gas ex Coal-gasification	CO_2, H_2S, HCN, org. S Hydrocarbons	Fischer-Tropsch Synthesis	39.600,000	Lurgi
1979	Germany	1	Raw Gas ex Oil-gasification	CO_2	Methanol Synthesis	1,000,000	Lurgi
1981	Venezuela	2	Raw Gas ex Oil-gasification	CO_2, H_2S, COS, HCN	Hydrogen	7,500,000	Lurgi
1981	Germany	2 1	Raw Gas ex Oil-gasification	H_2S COS, HCN, CO_2	Hydrogen, Methanol Synthesis	7,200,000/ 3,400,000	Lurgi
1983	Germany	1	Raw Gas ex Coal-gasification	H_2S COS, HCN, CO_2	Hydrogen+ Oxogas	1.870.000	Lurgi

STATUS SUMMARY

BENFIELD ACID GAS REMOVAL

1.0 GENERAL INFORMATION

Developer:
Benfield Corporation
Pittsburgh, Pa. 15228

Licensor:
Union Carbide Corporation
Engineering Product and Processes Department
P. O. Box 44
Tonawanda, New York 14150

Type: A chemical absorption process which uses a hot potassium carbonate solution as the solvent for removal of acid gases (CO_2, H_2S and COS) and other impurities from a variety of feed streams including synthesis gases produced by coal gasification and partial oxidation processes.

Conditions: Gas stream pressures in Benfield units range from under 100 psia to 2000 psia with acid gas content of the feed ranging from 5 to 50 vol%. The temperature of the rich solution leaving the bottom of the absorber is normally between 250 - 260°F.

Applications: The process has been applied to purifying gas streams for the following applications:

- Ammonia Synthesis Gas
- Hydrogen Streams
- Natural Gas for Pipeline or LNG.
- SNG based on liquid feedstock.
- Coal Gasification streams for High or Low BTU Gas.
- Partial Oxidation Gases from the Texaco or Shell Processes.
- Reducing Gases for Iron Ore Reduction.
- Ethylene Oxide, Methanol, Vinyl Acetate and Oxo-Synthesis Gases.

Process Scheme: There are four basic process schemes: a) single-stage design; b) split-flow absorber design; c) two-stage design and d) the HiPure process. For typical coal-to-SNG applications, the HiPure process is often used. Recent improvements to the Benfield process were made by Union Carbide to reduce energy consumption via the LoHeat Process. The design features of LoHeat Process can be incorporated into any one of the above four designs to reduce energy consumption. (Refer to Section 4.0 for details).

Status: Over 520 units have been designed and/or operated throughout the world including the installation at Westfield, Scotland, which purifies the raw gas from a Lurgi gasifier.

2.0 PROCESS DEVELOPMENT

- The Hot Potassium Carbonate Process was developed by H. E. Benson, J. H. Field and other co-workers at the U. S. Bureau of Mines, Bruceton, Pa., in the early 1950's.
- The process was developed with the objective of purifying synthesis gas produced from coal gasification resulting in a gas with high CO_2 partial pressure at high temperature and containing H_2S and COS.
- Benson and Field later formed the Benfield Corporation and licensed the process under the name of "Benfield Process"
- As the use of Hot Potassium Carbonate process expanded, catalyst activators were developed which increased rates of absorption and regeneration by several process licensors: Activated Hot Carbonate Process (Benfield), Catacarb (Eickmeyer & Associates); Giamarco-Vetrocoke (Giamarco-Vetrocoke, Italy).
- In the mid 1960's, the Power Gas Corporation (now Davy-McKee Ltd.) patented an adaptation of the Hot Potassium Carbonate process employing certain independent countercurrent solution circuits and reuse of stripping steam. This development and similar design are presently offered as the Benfield HiPure Process.
- Latest improvement to the Hot Potassium Carbonate processes is the reduction in thermal energy requirements via the Benfield LoHeat Process.

3.0 SOLVENT CHARACTERISTICS

- CO_2 and H_2S Absorption

The absorption-regeneration reactions of H_2S and CO_2 with activated hot potassium carbonate solution can be represented as:

$$K_2CO_3 + H_2S \longrightarrow KHCO_3 + KHS$$

$$K_2CO_3 + CO_2 + H_2O \longrightarrow 2KHCO_3$$

Equilibrium pressures of H_2S and CO_2 over the solution increase with temperature and concentrations of KHS and $KHCO_3$. Because $KHCO_3$ is formed when either CO_2 or H_2S is absorbed, their equilibrium pressures over a solution are interdependent and must be carefully considered. When the CO_2:H_2S ratio of the feed gas is greater than 8, a reasonable estimate of capital cost and utility requirements can be made by assuming total acid gas to be only CO_2. Since H_2S is more soluble than CO_2, its content in the purified gas will be reduced to a greater extent than that of CO_2, and thus nearly complete removal of H_2S can be obtained.

- Selective H_2S Absorption

Since H_2S is more soluble in the activated hot carbonate solution than is CO_2, H_2S is absorbed and stripped at a faster rate. Operating conditions can be selected such that an H_2S absorption rate 50 times that of CO_2 can be achieved. Generally, H_2S removal of 90 to 98% is possible while CO_2 removal can be restricted to 10 to 40%.

- COS, CS_2, Mercaptans and Thiophene Absorption.

COS hydrolyzes in activated hot potassium carbonate solution:

$$COS + H_2O \longrightarrow H_2S + CO_2$$

The reaction rate is slower than that for CO_2 absorption, but the reaction will proceed to completion. The rate of hydrolysis is only slightly dependent on solution conversion to $KHCO_3$ or KHS but is highly dependent on temperature. COS removal to 1 ppm level can be achieved in a slightly different process configuration for systems optimized for CO_2 and H_2S removal. In the presence of activated hot potassium carbonate solution, the CS_2 is hydrolyzed in two stages:

$$CS_2 + H_2O \longrightarrow COS + H_2S$$

$$COS + H_2O \longrightarrow CO_2 + H_2S$$

Since an additional step is involved, CS_2 is absorbed at a slower rate than COS.

Mercaptans are slightly acidic and react with activated hot carbonate solution as follows:

$$K_2CO_3 + RSH \longrightarrow RSK + KHCO_3$$

Absorption is rapid but proceeds only to a limited extent.

Thiopene would not be expected to react chemically with activated hot potassium carbonate solution.

- Ammonia and HCN Removal

  The absorption rate of NH_3 in activated hot potassium carbonate solution is high, but the solution capacity is low. The degree of removal is dependent on the concentration of ammonia in the feed gas and the solution circulation rate.

  HCN reacts readily with activated hot potassium carbonate solution:

  $$K_2CO_3 + HCN \longrightarrow KCN + KHCO_3$$

  Removal efficiencies are expected to be as high as that for H_2S.

- Heavy hydrocarbons do not react with or degrade activated hot potassium carbonate solution. However, hydrocarbons condensates should be avoided since they can initiate foaming.

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- Relative absorption capacity and rate of absorption into Activated Hot Carbonate solution at 230°F is given below: (Ref. 2).

Component	Relative Capacity	Relative Rate
CO_2	1	1
H_2S	1.41	3.6
COS	Hydrolyze	.36
CS_2	Hydrolyze	.10
CH_3SH	0.03	1.2
NH_3	0.10	3.5

Capacities measured to an equilibrium partial pressure of 2 psi. Rates measured at solution loading equivalent to an equilibrium partial pressure of 2 psi.

4.0 PROCESS DESCRIPTION

The Benfield Activated Hot Potassium Carbonate Process employs conventional packed or tray towers. A flow diagram of a single stage process is shown in Figure 4-1. Both the absorber and regenerator operate isothermally at a temperature near the solution's atmospheric boiling point, normally about 230°F. The process gas to be purified is introduced at the bottom of the absorber and is contacted by lean solution flowing countercurrently downwards. Rich solution is withdrawn at the base of the absorber and is regenerated while passing downward through the second tower against a rising flow of steam. Stripping steam is usually supplied from a process gas or steam heated solution reboiler. The regenerated or lean solution is then recycled back to the top of the absorber. The CO_2 level of the purified gas can be reduced down to 1% with this process configuration.

A number of process variations are available to achieve higher degrees of purification, to satisfy specific process requirements.

- Split Flow Process (Figure 4-2) - In this process configuration, the bulk of the hot regenerated solution is introduced to an intermediate point of the absorber. A portion of this solution is cooled down to between 160 and 200°F and fed to the top of the tower. The lowered equilibrium pressure of CO_2/H_2S over the cooled solution allows a much higher gas purity to be achieved. The CO_2 content of the purified gas can be reduced down to 0.1%. The solution is regenerated in a conventional single column.

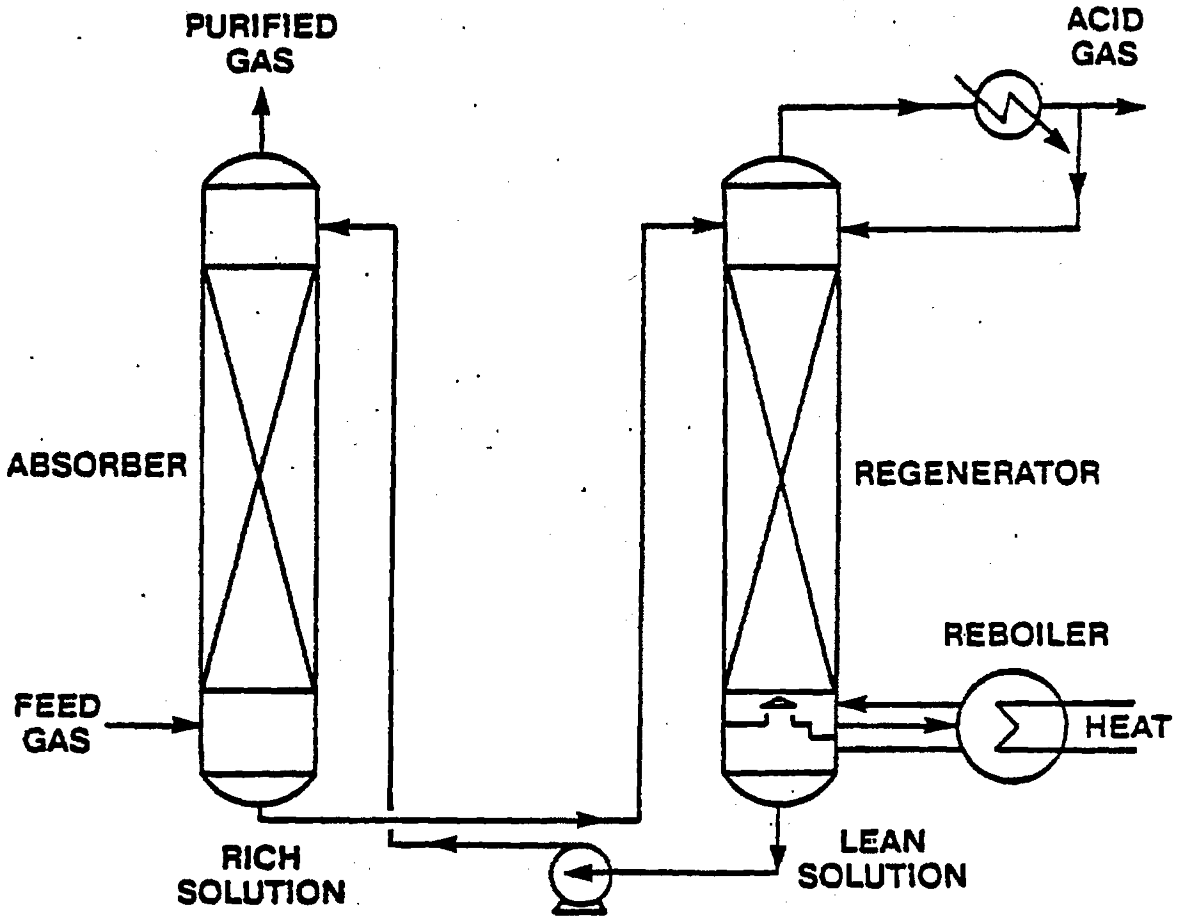
- Two-Stage Process (Figure 4-3) - In this system, both the absorber and the regenerator are split into two distinct sections. A major portion of the regenerated solution flow (semi-lean solvent) is split off from the mid-point of the absorber. The remainder is stripped more thoroughly with the total steam flow. This lean solution is cooled and then fed to the top of the absorber. The unit can be designed to achieve a typical CO_2 concentration of 500 ppm in the purified gas.

- Benfield Hi-Pure Process (Figure 4-4) - This process uses two independent circulating solutions in series to achieve a high purity product gas. The process gas is first contacted with normal Benfield hot carbonate solution in the primary absorber followed by contact with an amine solution in the secondary absorber. Bulk acid gas removal occurs in the bottom section of the primary absorber employing Benfield activated hot carbonate solution. Final acid gas cleanup occurs in the top section of the secondary absorber employing amine in a separate circuit at a reduced temperatures. Regeneration of the amine occurs in the bottom section of the secondary regenerator while the activated Benfield solution is regenerated in the top section of the

FIGURE 4-1

(SOURCE: REF. 1)

BENFIELD PROCESS SINGLE STAGE ABSORBER TYPICAL PURITY TO 1% CO_2



TYPICAL APPLICATIONS

Natural Gas

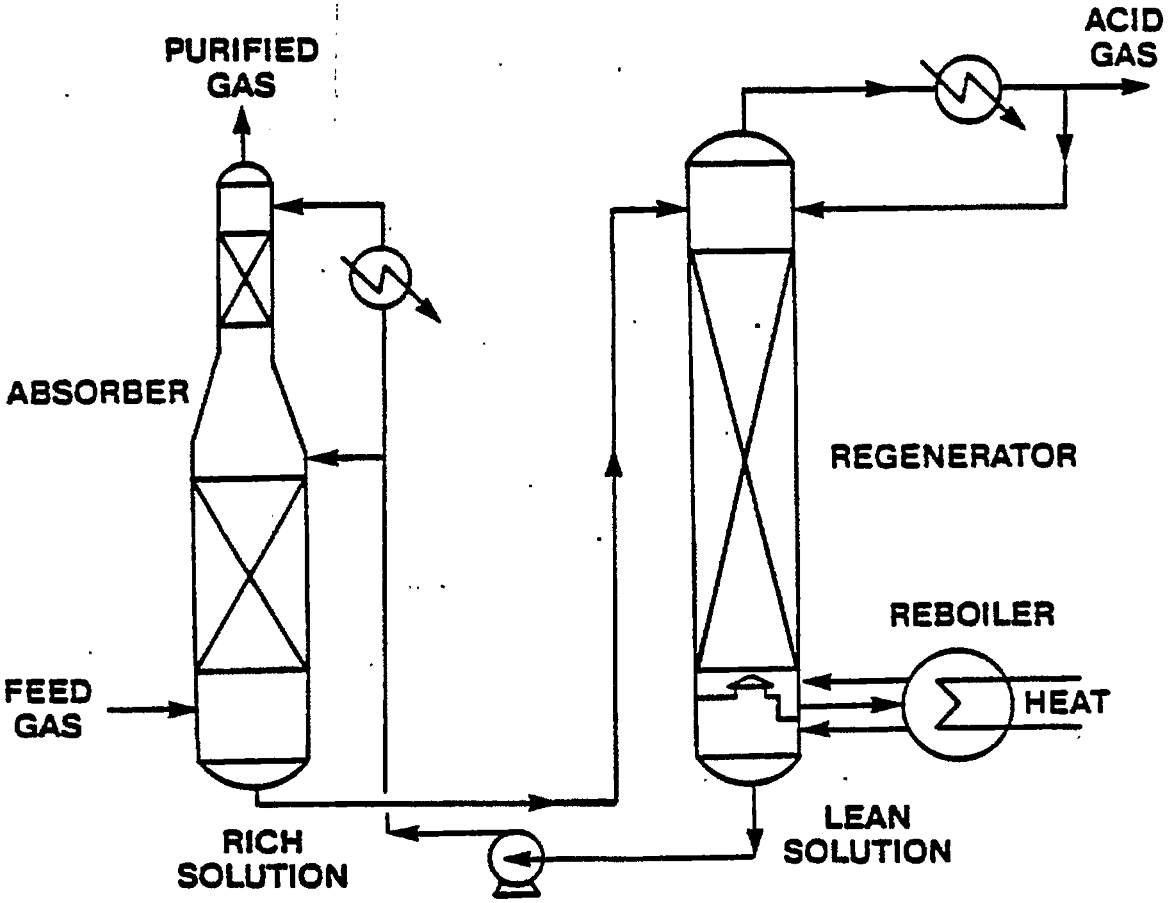
SNG

Ethylene Oxide

FIGURE 4-2

(SOURCE: REF. 1)

BENFIELD PROCESS SPLIT FLOW ABSORBER TYPICAL PURITY TO 0.1% CO_2



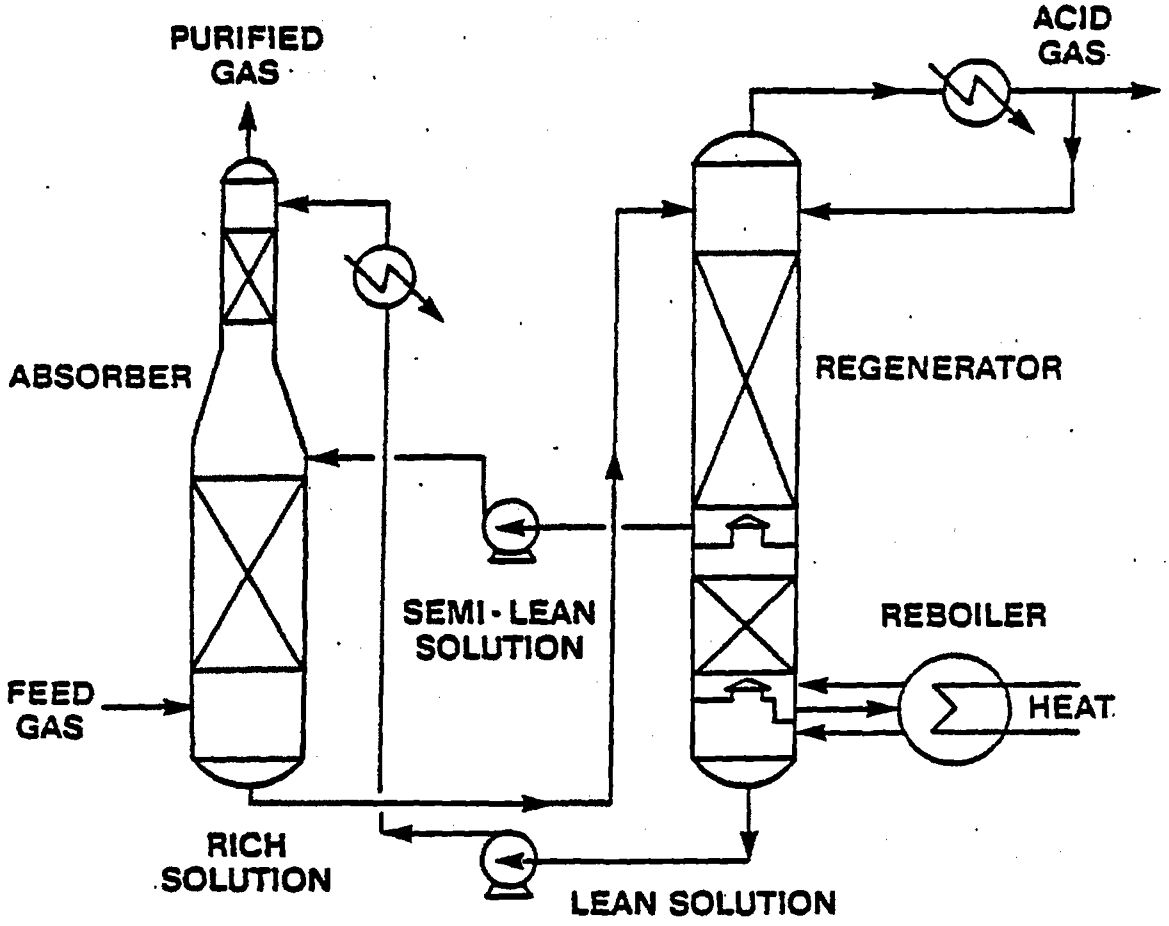
TYPICAL APPLICATIONS

- Ammonia Synthesis Gas
- Hydrogen
- Natural Gas

FIGURE 4-3

(SOURCE: REF. 1)

BENFIELD PROCESS TWO STAGE REGENERATOR TYPICAL PURITY TO 0.05% CO_2



TYPICAL APPLICATIONS

Ammonia Synthesis Gas

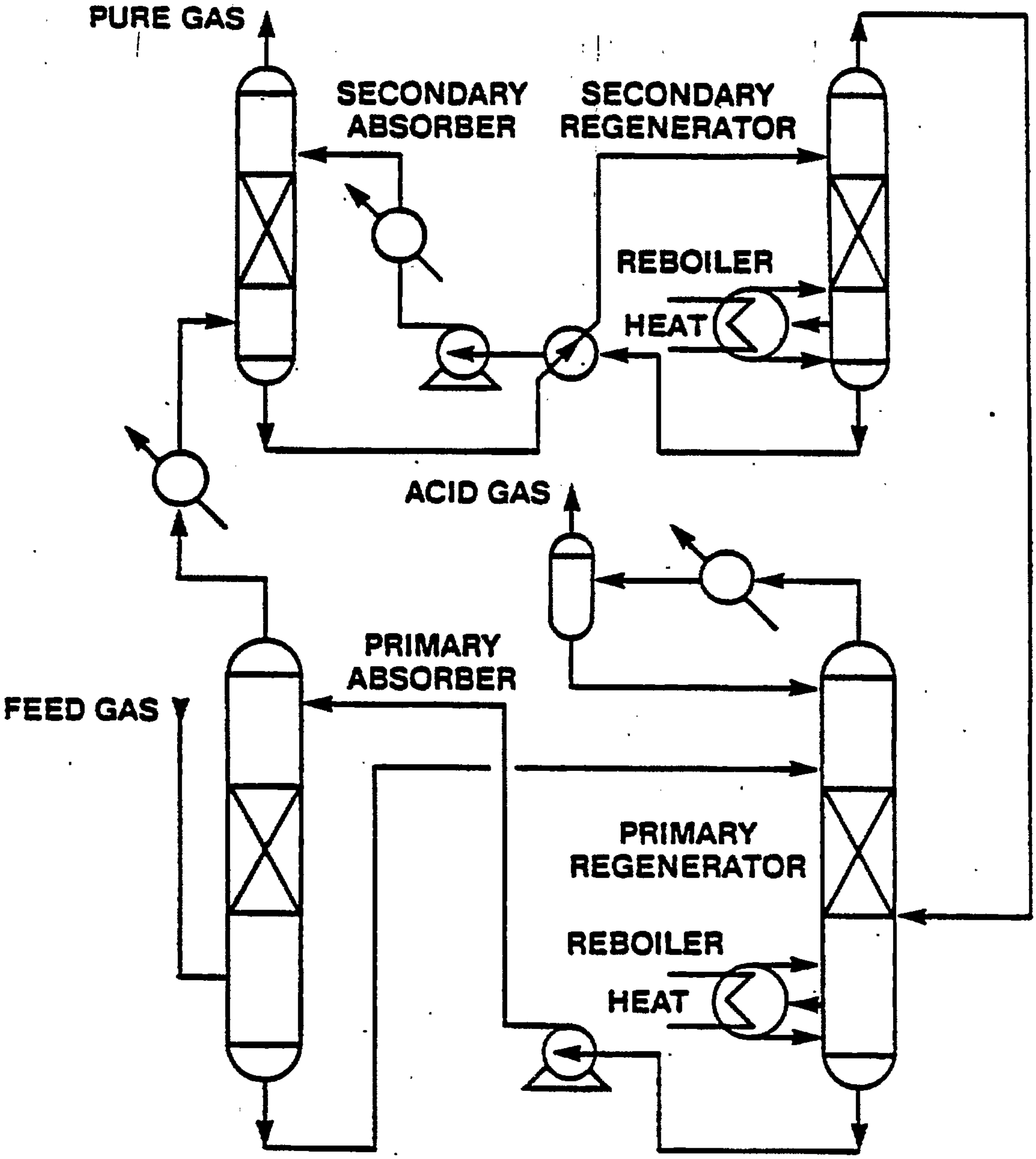
Hydrogen

Coal Gasification

FIGURE 4-4
(SOURCE: REF. 1)

BENFIELD HiPURE PROCESS UNIT

TYPICAL PURITIES TO < 50 PPM CO_2
< 1 PPM H_2S



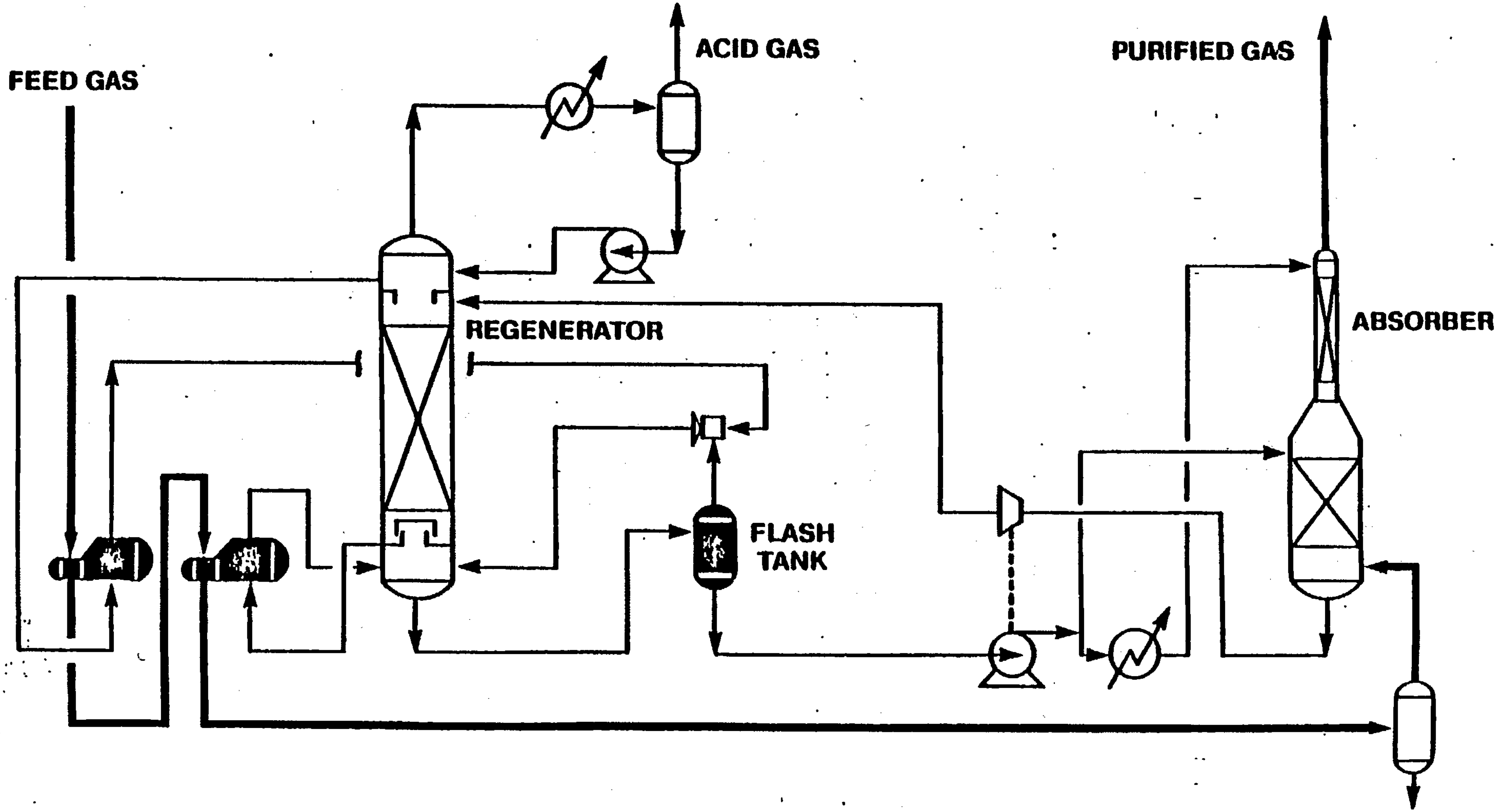
TYPICAL APPLICATIONS

Natural Gas for Pipeline
LNG Pretreatment

FIGURE 4-5

(SOURCE: REF. 1)

BENFIELD LOHEAT PROCESS
FOR CO_2 REMOVAL
UTILIZING STEAM EJECTORS



primary regenerator. Stripping steam used to thoroughly regenerate the solution flowing in the amine circuit is reused as stripping steam in the top portion of the regenerator. Purified gas with 1 ppm H_2S and 50 ppm CO_2 can be obtained with this process.

- Recent improvements to the Benfield Processes were made to reduce the regeneration energy consumption via the use of the LoHeat Process. In the LoHeat design, the regenerated lean solution is flashed at a lower pressure. The steam flashed off is compressed and used directly in the reboiler to strip the CO_2 and H_2S from the rich solution. Either thermocom-pressors (steam ejectors) or mechanical compressors can be used (See Figure 4-5). The net effects are:
  - Less outside heat required for the same regeneration and the same product gas purity;
  - Required reboiler surface area is decreased, and
  - The acid gas and lean solution cooling duties are decreased.

5.0 COMMERCIAL DESIGN DATA

A set of commercial Benfield HiPure acid gas removal system design data for a 250 Billion Btu/day coal-to-SNG plant is shown in Table 5-1. The feed gas is from the gasification of N. Dakota lignite in the Westinghouse gasifiers. Table 5-2 gives the calculation of contribution to gas cost based on design information given in Table 5-1.

TABLE 5-1

DESIGN DATA FOR A BENFIELD AGR SYSTEM IN A 250 BILLION BTU/DAY COAL-TO-SNG PLANT

Stream *	Feed gas	Acid gas	Sweetened Gas
Flowrate, Moles/hr	71,126	56,606	43,576
Pressure, psia	406.0	15.7	404.0
Temp,deg F	180.0	122.0	122.0
Composition (Vol %)			
CO	13.35	0.06	21.71
H2	41.38	0.25	67.22
CH4	5.97	0.03	9.71
C2H6			
N2/AR	0.46	0.00	0.75
CO2	36.60	45.98	0.05
COS	0.03	0.00	1 ppm
H2S	0.36	0.48	4 ppm
H2O	1.85	53.19	0.57

UTILITY REQUIREMENTS

Electric Power	10,500 Kw
Steam	
60 psig	795,000 #/hr
Cooling Water	42,000 Gpm
Make-up Solvent	51 Gal/hr

CAPITAL COSTS

Equipment (Installed)	$ 48.5 MM (Mid-1982)
Royalties	$ 2.9 MM (Mid-1982)

* Refer to Figure 4-5

TABLE 5-2

CALCULATION OF CONTRIBUTION TO GAS COST
BENFIELD (NON-SELECTIVE) SYSTEM

Coal Type	N.Dakota lignite (Westinghouse Gasifiers)
Evaluator	Kellogg Rust Synfuels,Inc.
Project Report No.	None
Date Published	None
Plant Capacity	250 Billion Btu/day SNG

CAPITAL COSTS :	$ MM (Mid-1982)
Installed Equipment	48.5
Contingency @ 15%	7.3
Direct Facility Constr Investment	55.8
Home-Office costs @ 12%	6.7
Total Facility Constr Investment	62.5
Royalties	2.9
Initial Solvent Charge	2.4
Total Plant Investment	67.8

OPERATING COSTS :			$/hr
Steam(60 psig)	795,000 #/hr	@ $ 3.85/ 1000 lb.	3060.8
Electricity	10,500 Kw	@ $0.05/Kwh	525.0
Cooling water	42,000 Gpm	@ $0.10/ 1000 Gal	252.0
Solvent Make-up	51 Gal/hr	@ $2.00/Gal	102.0
TOTAL			3939.8

Total Operating Cost, $ MM/yr at 100 % Stream factor = 34.5 MM $/Yr

CONTRIBUTION TO GAS COSTS :	Specific Cost, $/MM Btu-Yr	Charge Rate, Year	Contribution, $/MM Btu
Capital Related	0.83	0.089	0.07
Operating	0.42	1.000	0.42
Total			0.49

6.0 ADVANTAGES & DISADVANTAGES

- Advantages
  - Low cost non-proprietary chemicals are used.
  - The solution is stable, hence no solution reclaimers are required and no fouling of heat exchangers and column internals occurs.
  - Rich/lean solution heat exchangers are usually eliminated.
  - Process gas coolers before acid gas removal are sometimes eliminated because hot gases can be treated. This results in improving the overall thermal efficiency, although the regeneration steam requirement is relatively high as compared to other AGR processes. Low solubility of process gases minimizes product losses.
  - Most of the equipment can be of carbon steel construction.
  - A COS hydrolysis step is not required since the majority of COS is hydrolyzed to H_2S in the hot carbonate solution.

- Disadvantages
  - The Benfield process generates a very low H_2S concentration in the acid off-gas stream to the sulfur plant, hence requiring a more expensive sulfur recovery facility.
  - When the CO content of the feed gas is high, or at high temperature, CO will react with K_2CO_3 to form potassium formate (KCaOH). A small purge stream is necessary to remove the formate and fresh carbonate solution makeup is required.

- Hydrocarbon Losses: (Typical) (Ref: 3, 4, 5)

Component	Losses, as percent of Feed
H_2	<1%
CO	<1%
CH_4	<1%
C_2H_6	<1%

7.0 COMMERCIAL INSTALLATIONS

A list of all the Benfield Activated Hot Carbonate Process Installation is shown in Table 7-1. (Source: Ref. 3)

8.0 REFERENCES

1. Bartoo, R. K., "The Benfield Process for Acid Gas Removal". Paper presented at the AIChE 1984 Winter National Meeting, March 1984, Atlanta, Georgia.

2. Parrish, R. W., and H. B. Neilsen, "Synthesis Gas purification Including Removal of Trace Contaminants by the Benfield Process". Paper presented at the 167th National ACS Meeting. Division of Industrial and Engineering Chemistry, Los Angeles, California, 1974.

3. Personal Communication with Union Carbide Corp. Engineering Products & Processes Division.

4. Cassano, A. A. et al., "Cryogenic Methane Separation/Catalytic Hydrgrasification Process Analysis", Final Report, February 1980, FE-3044-T12.

5. Kellogg In-House data.

TABLE 7-1

(SOURCE: REF. 3)

BENFIELD CORPORATION
615 WASHINGTON ROAD
PITTSBURGH, PA 15228

GENERAL PLANT LIST

	PLANT OWNER	LOCATION	CONTRACTOR/ENGINEER	TYPE	*SIZE	CO2 IN	CO2 OUT	H2S IN	H2S OUT	NOTE
1)	LONE STAR GAS CO.	TEXAS, USA		NAT. GAS	92.0 MM	8.5%	0.500%	1.500%	4000 PPM	(1)
2)	PETROGAS PROCESSING	CANADA	RALPH M PARSONS CO	NAT. GAS	125.0 MM	7.0%	1.000%	16.000%	1000 PPM	(1)
3)	HELWAN AMMONIA	EGYPT	DIDIER ENGRG GMBH	AMMONIA	190 TPD	14.5%	0.500%	--	--	(1)
4)	HELWAN AMMONIA	EGYPT	DIDIER ENGRG GMBH	AMMONIA	190 TPD	14.5%	0.500%	--	--	(1)
5)	BROCKVILLE CHEMICAL	CANADA	DAVY POWERGAS	AMMONIA	250 TPD	24.9%	0.400%	--	--	(1)
6)	DOW CHEMICAL CO.	TEXAS, USA	PACE ENGINEERING	CONFID.	7.7 MM	10.0%	0.500%	--	--	(1)
7)	CHEMOKOMPLEX	HUNGARY	SIMON CARVES	AMMONIA	495 TPD	22.0%	0.100%	--	--	(1)
8)	POLIMEX	POLAND	SIMON CARVES	AMMONIA	550 TPD	18.0%	0.100%	--	--	(1)
9)	POLIMEX	POLAND	SIMON CARVES	AMMONIA	550 TPD	18.0%	0.100%	--	--	(1)
10)	POLIMEX	POLAND	SIMON CARVES	AMMONIA	550 TPD	18.0%	0.100%	--	--	(1)
11)	IMPERIAL CHEM INDUST	GREAT BRITAIN	DAVY POWERGAS	AMMONIA	300 TPD	19.8%	0.070%	--	--	(1)
12)	PEMEX	MEXICO	LUMMUS COMPANY	AMMONIA	200 TPD	16.6%	0.500%	--	--	(1)
13)	E. I. DUPONT	WEST VIR., USA	PLANT OWNER	AMMONIA	OPERATING DETAILS ARE CONFIDENTIAL					(1)
14)	APPLE RIVER CHEMICAL	ILLINOIS, USA	LUMMUS COMPANY	AMMONIA	600 TPD	18.1%	0.100%	--	--	(1)
15)	STAVELEY CHEMICAL	GREAT BRITAIN	SELAS B. V.	HYDROGEN	6.0 MM	23.9%	0.300%	--	--	(1)
16)	DELHI CLOTH MILLS	INDIA	CHIYODA CHEM ENGR	AMMONIA	495 TPD	22.0%	0.100%	--	--	(1)
17)	A. P. C. (P. E. C. -RHIN)	FRANCE	PULLMAN KELLOGG	AMMONIA	120 TPD	15.5%	0.400%	--	--	(1)
18)	A. P. C. (P. E. C. -RHIN)	FRANCE	PULLMAN KELLOGG	AMMONIA	120 TPD	15.5%	0.400%	--	--	(1)
19)	ICI MOND DIVISION	GREAT BRITAIN	DAVY POWERGAS	CARB MON	4.9 MM	23.1%	0.500%	--	--	(1)
20)	IDEMITSU KOSAN	JAPAN	FRIEDRICH UHDE	HYDROGEN	33.0 MM	24.2%	0.100%	--	--	(1)
21)	S. N. A. M.	ITALY	WOODALL DUCKHAM	TOWN GAS	53.0 MM	26.3%	5.400%	--	--	(1)
22)	S. N. A. M.	ITALY	WOODALL DUCKHAM	TOWN GAS	53.0 MM	26.3%	5.400%	--	--	(1)
23)	RIO GULF	SPAIN	LUMMUS COMPANY	HYDROGEN	23.0 MM	16.0%	0.200%	--	--	(1)
24)	PETROLEOS MEXICANOS	MEXICO	LUMMUS COMPANY	AMMONIA	1000 TPD	18.1%	0.100%	--	--	(1)
25)	E. I DUPONT	TEXAS, USA	PLANT OWNER	AMMONIA	OPERATING DETAILS ARE CONFIDENTIAL					(1)
26)	SYRIAN GOVERNMENT	SYRIA	SELAS B. V.	HYDROGEN	14.0 MM	23.3%	0.200%	--	--	(1)
27)	ULTRAFERTIL S. A.	BRAZIL	FOSTER WHEELER CORP	AMMONIA	500 TPD	22.0%	0.050%	--	--	(1)
28)	S. E. I. F. NANGIS	FRANCE	SOC. TECH. ENT. CHIM.	AMMONIA	1000 TPD	22.0%	0.100%	--	--	(1)
29)	POLIMEX	POLAND	HALDOR TOPSOE	AMMONIA	825 TPD	17.9%	0.100%	--	--	(1)
30)	POLIMEX	POLAND	HALDOR TOPSOE	AMMONIA	825 TPD	17.9%	0.100%	---	--	(1)
31)	PETROLEOS MEXICANOS	MEXICO	BECHTEL CORP	HYDROGEN	20.0 MM	18.6%	0.100%	--	--	(1)
32)	TYPPI OY	FINLAND	HUMPHREYS & GLASGOW	AMMONIA	750 TPD	22.0%	0.100%	--	--	(1)
33)	NIPPON PETROLEUM	JAPAN	DAVY POWERGAS	HYDROGEN	21.6 MM	22.8%	0.100%	--	--	(2)
34)	TOA NENRYO KOGYO KK	JAPAN	CHIYODA CHEM ENGR	HYDROGEN	15.6 MM	21.5%	0.100%	--	--	(1)
35)	DAIKYO OIL COMPANY	JAPAN	FOSTER WHEELER CORP	HYDROGEN	11.6 MM	23.1%	0.100%	--	--	(1)
36)	MARUZEN SEKIYU CO.	JAPAN	CHIYODA CHEM ENGR	HYDROGEN	21.6 MM	22.9%	0.100%	--	--	(1)
37)	MO OCH DOMSJO	SWEDEN	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
38)	A. P. C. (P. E. C. -RHIN)	FRANCE	FRIEDRICH UHDE	AMMONIA	545 TPD	21.8%	0.100%	--	--	(1)
39)	I. E. L - KANPUR	INDIA	IMP. CHEM. IND. (ICI)	AMMONIA	457 TPD	22.7%	0.200%	--	--	(1)
40)	I. E. L - KANPUR	INDIA	IMP. CHEM. IND. (ICI)	AMMONIA	457 TPD	22.7%	0.200%	--	--	(1)
41)	NATURAL GAS OF N. Z.	NEW ZEALAND	SCIEN. DESIGN LTD	NAT. GAS	15.8 MM	50.0%	3.000%	--	--	(1)
42)	NATURAL GAS OF N. Z.	NEW ZEALAND	SCIEN. DESIGN LTD	NAT. GAS	15.8 MM	50.0%	3.000%	--	--	(1)

* - SIZE IS EITHER IN MILLIONS OF NORMAL CUBIC FEET OF PRODUCT GAS PER DAY OR IN TONS OF PRODUCT PER DAY

NOTES - (1) CONVENTIONAL BENFIELD (2) HIPURE (3) CATACARB CONVERSION (4) VETROCOKE CONVERSION (5) OTHER CONVERSION

	PLANT OWNER	LOCATION	CONTRACTOR/ENGINEER	TYPE	*SIZE	CO2 IN	CO2 OUT	H2S IN	H2S OUT	NOTE
43)	FUJI OIL COMPANY	JAPAN	CHIYODA CHEM ENGR	HYDROGEN	11.7 MM	24.9%	0.100%	--	--	(1)
44)	GEWERKSCHAFT VICTOR	WEST GERMANY	FRIEDRICH UHDE	AMMONIA	1000 TPD	18.3%	0.100%	--	--	(1)
45)	KOA OIL COMPANY LTD	JAPAN	CHIYODA CHEM ENGR	HYDROGEN	6.2 MM	24.7%	0.100%	--	--	(1)
46)	TECHNOIMPORT	BULGARIA	HUMPHREYS & GLASGOW	AMMONIA	620 TPD	22.3%	0.100%	--	--	(1)
47)	KASHIMA OIL COMPANY	JAPAN	DAVY POWERGAS	HYDROGEN	2.4 MM	29.8%	1.300%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	2.3 MM	8.3%	0.090%	--	--	
48)	KASHIMA OIL COMPANY	JAPAN	DAVY POWERGAS	HYDROGEN	9.6 MM	5.3%	0.600%	0.630%	10 PPM	(2)
49)	PPG INDUSTRIES	PUERTO RICO	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
50)	NORTHERN PETROCHEM	ILLINOIS, USA	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
51)	NISSO PETROCHEMICALS	JAPAN	SCIENTIFIC DESIGN	NAT. GAS	105.5 MM	7.2%	0.250%	0.002%	3 PPM	(1)
52)	TERNI	ITALY	FOSTER WHEELER CORP	AMMONIA	225 TPD	17.9%	0.100%	--	--	(1)
53)	DUTCH STATE MINES	NETHERLANDS	BECHTEL CORP	AMMONIA	1500 TPD	17.8%	0.150%	--	--	(1)
54)	VEB SYNTHESE WERKE	EAST GERMANY	SPEICHIM - ONIA GEGI	HYDROGEN	4.8 MM	22.8%	0.070%	--	--	(1)
55)	INTERNATIONAL NICKEL	GREAT BRITAIN	DAVY POWERGAS	HYDROGEN	2.3 MM	24.3%	0.100%	--	--	(2)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			CARB MON	4.9 MM	15.2%	0.100%	--	--	
56)	ASAHI CHEMICAL INDUS	JAPAN	C. F. BRAUN	AMMONIA	1000 TPD	16.8%	0.050%	--	--	(1)
57)	CSR CHEMICALS LTD	AUSTRALIA	FRIEDRICH UHDE	VIN. AC.	0.5 MM	19.9%	2.000%	--	--	(1)
58)	TAESUNG LUMBER CO	SOUTH KOREA	DAVY POWERGAS	METHANOL	0.4 MM	12,4%	1.500%	--	--	(1)
59)	BRITISH PETROLEUM	FRANCE	FOSTER WHEELER CORP	HYDROGEN	28.0 MM	21.1%	0.100%	--	--	(1)
60)	NISSON PETROCHEMICAL	JAPAN	HITACHI LIMITED	OXO GAS	3.8 MM	25.9%	0.500%	--	--	(1)
61)	WEST COAST TRANSMISS	CANADA	PLANT OWNER	NAT. GAS	165.0 MM	12.3%	0.800%	0.500%	10 PPM	(5)
62)	YACIM. PETRO. FISCAL.	ARGENTINA	LUMMUS COMPANY	HYDROGEN	18.4 MM	20.4%	0.150%	--	--	(1)
63)	KANSAI OIL COMPANY	JAPAN	CHIYODA CHEM ENGR	HYDROGEN	10.6 MM	24.6%	0.200%	--	--	(1)
64)	PETROCHEMICAL INDUST	KUWAIT	HALDOR TOPSOE	AMMONIA	875 TPD	19.7%	0.100%	--	--	(1)
65)	PETROCHEMICAL INDUST	KUWAIT	HALDOR TOPSOE	AMMONIA	875 TPD	19.7%	0.100%	--	--	(1)
66)	DOW CHEMIE N. V.	NETHERLANDS	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
67)	ZDRUZENIE CHEMIOKEHO	CZECHOSLOVAKIA	FRIEDRICH UHDE	AMMONIA	1000 TPD	18.0%	0.100%	--	--	(1)
68)	BP CHEMICALS LTD	GREAT BRITAIN	LUMMUS COMPANY	VIN. AC.	1.9 MM	27.5%	8.600%	--	--	(1)
69)	KYOWA YUKA KK	JAPAN	HITACHI LIMITED	OXO GAS	8.6 MM	40.4%	0.100%	--	--	(1)
70)	INDUSTRIAL IMPORT	ROMANIA	FOSTER WHEELER CORP	AMMONIA	1000 TPD	18.6%	0.100%	--	--	(1)
71)	PETROLEOS MEXICANOS	MEXICO	BECHTEL CORP	HYDROGEN	51.0 MM	19.8%	0.100%	--	--	(1)
72)	QATAR FERTILIZER	QATAR	DAVY POWERGAS	NAT. GAS	17.1 MM	7.0%	0.180%	1.000%	4 PPM	(1)
73)	QATAR FERTILIZER	QATAR	DAVY POWERGAS	AMMONIA	1000 TPD	18.8%	0.120%	--	--	(1)
74)	E. I. DUPONT	TEXAS, USA	FRIEDRICH UHDE	VIN. AC.	OPERATING DETAILS ARE CONFIDENTIAL					(1)
75)	IDEMITSU KOSAN	JAPAN	FOSTER WHEELER CORP	HYDROGEN	16.4 MM	24.9%	0.100%	--	--	(1)
76)	GENERAL OIL COMPANY	JAPAN	DAVY POWERGAS	HYDROGEN	24.0 MM	23.4%	0.100%	--	--	(2)
77)	KASEI MIZUSHIMA CO	JAPAN	FRIEDRICH UHDE	HYDROGEN	10.5 MM	19.0%	0.200%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			OXO GAS	8.6 MM	27.8%	1.000%	--	--	
78)	OSAKA GAS COMPANY	JAPAN	HUMPHREYS & GLASGOW	TOWN GAS	1.4 MM	21.2%	0.860%	--	--	(1)
79)	TOA NENRYO	JAPAN	CHIYODA CHEM ENGR	HYDROGEN	20.4 MM	23.9%	0.300%	--	--	(1)
80)	GULF SEKIYU SEISEI	OKINAWA	CHIYODA CHEM ENGR	HYDROGEN	44.1 MM	24.3%	0.200%	--	--	(1)
81)	NATIONAL DISTILLERS	TEXAS, USA	PLANT OWNER	VIN. AC.	12.0 MM	5.0%	2.000%	--	--	(5)
82)	POLIMEX	POLAND	HALDOR TOPSOE	AMMONIA	500 TPD	23.8%	0.430%	--	--	(1)
83)	POLIMEX	POLAND	HALDOR TOPSOE	AMMONIA	750 TPD	18.0%	0.100%	--	--	(1)
84)	AIR PROD (NETH) NV	NETHERLANDS	SELAS B. V.	HYDROGEN	17.0 MM	17.9%	0.100%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			CARB MON	1.1 MM	24.3%	0.100%	--	--	

* - SIZE IS EITHER IN MILLIONS OF NORMAL CUBIC FEET OF PRODUCT GAS PER DAY OR IN TONS OF PRODUCT PER DAY

NOTES - (1) CONVENTIONAL BENFIELD (2) HIPURE (3) CATACARB CONVERSION (4) VETROCOKE CONVERSION (5) OTHER CONVERSION

	PLANT OWNER	LOCATION	CONTRACTOR/ENGINEER	TYPE	*SIZE	CO2 IN	CO2 OUT	H2S IN	H2S OUT	NOTE
85)	NYPRO (U.K.) LTD	GREAT BRITAIN	DAVY POWERGAS	HYDROGEN	8.4 MM	19.2%	0.100%	--	--	(1)
86)	ORINOCO CHEMICAL	ITALY	BAMAG	AMMONIA	150 TPD	18.1%	0.080%	--	--	(1)
87)	GULF OIL COMPANY	CALIFORNIA, USA	PLANT OWNER	HYDROGEN	10.2 MM	21.1%	0.300%	--	--	(4)
88)	PETROLEOS MEXICANOS	MEXICO	BECHTEL CORP	HYDROGEN	51.0 MM	19.8%	0.100%	--	--	(1)
89)	KYOKUTO PETROL IND	JAPAN	UNIVERSAL OIL PROD	HYDROGEN	25.0 MM	24.7%	0.160%	--	--	(1)
90)	GARDINIER, INC.	GEORGIA, USA	PLANT OWNER	AMMONIA	400 TPD	15.0%	0.430%	--	--	(1)
91)	ASIA KYOSEKI CO LTD	JAPAN	CHIYODA CHEM ENGR	HYDROGEN	10.6 MM	22.9%	0.150%	--	--	(1)
92)	PETROFINA	CANADA	LUMMUS COMPANY	HYDROGEN	43.2 MM	16.8%	0.100%	--	--	(1)
93)	IDEMITSU KOSAN	JAPAN	CHIYODA CHEM ENGR	HYDROGEN	29.0 MM	24.3%	0.200%	--	--	(1)
94)	OSAKA GAS COMPANY	JAPAN	DAVY POWERGAS	TOWN GAS	5.1 MM	21.1%	3.000%	--	--	(1)
95)	OSAKA GAS COMPANY	JAPAN	DAVY POWERGAS	TOWN GAS	5.1 MM	21.1%	3.000%	--	--	(1)
96)	SHOWA YOKKAICHI	JAPAN	FOSTER WHEELER CORP	HYDROGEN	31.5 MM	24.1%	0.100%	--	--	(1)
97)	COLUMBIA NIPRO	GEORGIA, USA	DAVY POWERGAS	HYDROGEN	6.7 MM	19.2%	0.100%	--	--	(1)
98)	HUMBLE OIL CO	ALABAMA, USA	HUDSON ENGINEERING	NAT. GAS	14.7 MM	42.8%	2.000%	10.900%	3 PPM	(2)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			NAT. GAS	1.0 MM	29.0%	1.000%	20.000%	3 PPM	
99)	ICI NOBEL DIVISION	GREAT BRITAIN	DAVY POWERGAS	HYDROGEN	2.0 MM	23.5%	0.100%	--	--	(1)
100)	DELHI CLOTH MILLS	INDIA	CHIYODA CHEM ENGR	AMMONIA	150 TPD	22.0%	0.100%	--	--	(1)
101)	ABUDHABI GAS LIQ CO	DAS ISLAND	BECHTEL INT.	LNG	203.4 MM	4.9%	0.005%	4.720%	1 PPM	(2)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			NAT. GAS	57.4 MM	6.2%	0.620%	2.100%	300 PPM	
102)	ABUDHABI GAS LIQ CO	DAS ISLAND	BECHTEL INT.	LNG	203.4 MM	4.9%	0.005%	4.720%	1 PPM	(2)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			NAT. GAS	57.4 MM	6.2%	0.620%	2.100%	300 PPM	
103)	COLUMBIA LNG CORP.	OHIO, USA	DAVY POWERGAS	SNG	118.0 MM	17.9%	0.100%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	6.1 MM	21.6%	2.000%	--	--	
104)	COLUMBIA LNG CORP.	OHIO, USA	DAVY POWERGAS	SNG	118.0 MM	17.9%	0.100%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	6.1 MM	21.6%	2.000%	--	--	
105)	CONSUMERS POWER CO.	MICHIGAN, USA	LUMMUS COMPANY	SNG	103.8 MM	21.9%	1.000%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	12.0 MM	22.9%	0.300%	--	--	
106)	CONSUMERS POWER CO.	MICHIGAN, USA	LUMMUS COMPANY	SNG	103.8 MM	21.9%	1.000%	--	--	(1)
107)	NIPPON PETROLEUM	JAPAN	DAVY POWERGAS	HYDROGEN	20.7 MM	23.5%	0.200%	--	--	(1)
108)	ETHYLOX	FRANCE	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
109)	KASEI MIZUSHIMA	JAPAN	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
110)	SANTOS LTD TRAIN A	AUSTRALIA	GLOBAL ENGINEERING	NAT. GAS	109.1 MM	14.2%	2.500%	0.005%	1 PPM	(4)
111)	SANTOS LTD TRAIN B	AUSTRALIA	GLOBAL ENGINEERING	NAT. GAS	109.1 MM	14.2%	2.500%	0.005%	1 PPM	(4)
112)	SANTOS LTD TRAIN C	AUSTRALIA	GLOBAL ENGINEERING	NAT. GAS	116.1 MM	16.0%	2.000%	0.005%	1 PPM	(1)
113)	W. R. GRACE & CO	TENNESSEE, USA	FOSTER WHEELER CORP	AMMONIA	1000 TPD	18.2%	0.050%	--	--	(1)
114)	MANGALORE FERTILIZER	INDIA	HUMPHREYS & GLASGOW	AMMONIA	660 TPD	22.7%	0.150%	--	--	(1)
115)	BAHRAIN PETROLEUM	BAHRAIN	RALPH M PARSONS CO	HYDROGEN	26.4 MM	20.2%	0.150%	--	--	(1)
116)	NORTH. ILLINOIS GAS	ILLINOIS, USA	BECHTEL CORP	SNG	78.7 MM	21.0%	1.000%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	4.1 MM	20.0%	1.400%	--	--	
117)	NORTH. ILLINOIS GAS	ILLINOIS, USA	BECHTEL CORP	SNG	78.7 MM	21.0%	1.000%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	4.1 MM	20.0%	1.400%	--	--	
118)	IMP CHEM IND AUS&NZ	NEW ZEALAND	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
119)	OXITENO S.A.	BRAZIL	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
120)	MITSUI PETROCHEMICAL	JAPAN	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
121)	O. S. W.	AUSTRIA	FRIEDRICH UHDE	AMMONIA	850 TPD	21.7%	0.100%	--	--	(1)
122)	ROMCHIM	ROMANIA	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)

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NOTES - (1) CONVENTIONAL BENFIELD (2) HIPURE (3) CATACARB CONVERSION (4) VETROCOKE CONVERSION (5) OTHER CONVERSION

	PLANT OWNER	LOCATION	CONTRACTOR/ENGINEER	TYPE	*SIZE	CO2 IN	CO2 OUT	H2S IN	H2S OUT	NOTE
123)	MARINDUQUE MINING	PHILIPPINES	BECHTEL CORP	HYDROGEN	40.7 MM	22.4%	0.100%	--	--	(1)
124)	F.C.I. GORAKHPUR	INDIA	FERT. CORP. OF INDIA	AMMONIA	180 TPD	32.4%	2.000%	--	--	(1)
125)	ENTASA	SPAIN	UNIVERSAL OIL PROD	HYDROGEN	9.6 MM	19.4%	0 100%	--	--	(1)
126)	TOA OIL COMPANY	JAPAN	JAPAN GASOLINE	HYDROGEN	20.4 MM	13.8%	0.200%	--	--	(1)
127)	GREENVALE NICKEL	AUSTRALIA	RALPH M PARSONS CO	AMMONIA	55 TPD	21.8%	0.200%	--	--	(1)
128)	BROOKLYN UNION GAS	NEW YORK, USA	LUMMUS COMPANY	SNG	120.0 MM	23.7%	0.100%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	3.4 MM	23.3%	0.200%	--	--	
129)	ALGONQUIN GAS CO.	MASS., USA	LUMMUS COMPANY	SNG	120.0 MM	23.7%	0.100%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	6.8 MM	23.3%	0.200%	--	--	
130)	F.C.I. TROMBAY	INDIA	FERT. CORP. OF INDIA	AMMONIA	365 TPD	31.2%	0.800%	0.030%	10 PPM	(5)
131)	F.C.I. TROMBAY	INDIA	FERT. CORP. OF INDIA	AMMONIA	175 TPD	31.2%	0.800%	--	--	(1)
132)	F.C.I. TROMBAY	INDIA	FERT. CORP. OF INDIA	AMMONIA	175 TPD	31.2%	0.800%	--	--	(1)
133)	CHEVRON-SUN SACROC I	TEXAS, USA	OLSEN ENGINEERING	NAT. GAS	57.4 MM	24.0%	1.000%	0.002%	1 PPM	(1)
134)	CHEVRON-SUN SACROC I	TEXAS, USA	OLSEN ENGINEERING	NAT. GAS	57.4 MM	24.0%	1.000%	0.002%	1 PPM	(1)
135)	TOKYO GAS	JAPAN	DAVY POWERGAS	SNG	40.0 MM	23.7%	2.000%	--	--	(1)
136)	PEOPLES GAS COMPANY	ILLINOIS, USA	PULLMAN KELLOGG	HYDROGEN	9.1 MM	21.7%	1.000%	--	--	(1)
137)	PEOPLES GAS COMPANY	ILLINOIS, USA	PULLMAN KELLOGG	SNG	75.8 MM	22.3%	1.000%	--	--	(1)
138)	PEOPLES GAS COMPANY	ILLINOIS, USA	PULLMAN KELLOGG	SNG	75.8 MM	22.3%	1.000%	--	--	(1)
139)	ASHLAND OIL	NEW YORK, USA	FOSTER WHEELER CORP	SNG	54.9 MM	23.3%	2.500%	--	--	(1)
140)	PUBLIC SERVICE E&G	NEW JERSEY, USA	FOSTER WHEELER CORP	SNG	18.7 MM	22.5%	0.500%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	4.1 MM	20.0%	1.400%	--	--	
141)	PUBLIC SERVICE E&G	NEW JERSEY, USA	WOODALL DUCKHAM	SNG	122.2 MM	22.9%	0 600%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	6.1 MM	22.1%	0.500%	--	--	
142)	CHINESE PETROLEUM	TAIWAN	RALPH M PARSONS CO	HYDROGEN	48.6 MM	24.6%	0.100%	--	--	(1)
143)	CHISSO PETROLEUM	JAPAN	JAPAN GASOLINE	HYDROGEN	2.5 MM	46.2%	1.000%	--	--	(4)
144)	BARODA ATOM. RES. COM.	INDIA	HUMPHREYS & GLASGOW	AMMONIA	100 TPD	18.9%	0.100%	--	--	(1)
145)	INDIAN PETRO CHEM	INDIA	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
146)	GASCO	HAWAII, USA	RALPH M PARSONS CO	SNG	13.3 MM	24.2%	1.000%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	2.8 MM	23.0%	1.000%	--	--	
147)	GASCO	HAWAII, USA	RALPH M PARSONS CO	SNG	13.3 MM	24.2%	1.000%	--	--	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	2.8 MM	23.0%	1.000%	--	--	
148)	IDEMITSU KOSAN	JAPAN	FOSTER WHEELER CORP	HYDROGEN	19.2 MM	24.7%	0.200%	--	--	(1)
149)	KOCH REFINING CO	MINNESOTA, USA	TIELROOY & ASSOC	HYDROGEN	13.8 MM	24.4%	0.100%	--	--	(1)
150)	ISTANBUL GUBRE SAN.	TURKEY	FRIEDRICH UHDE	AMMONIA	1100 TPD	22.2%	0.100%	--	--	(1)
151)	CHEVRON OIL CO.	NEW JERSEY, USA	FOSTER WHEELER CORP	HYDROGEN	7.8 MM	22.6%	0.100%	--	--	(1)
152)	IDEMITSU KOSAN	JAPAN	FOSTER WHEELER CORP	HYDROGEN	15.0 MM	23.5%	0.200%	--	--	(1)
153)	IDEMITSU KOSAN	JAPAN	CHIYODA CHEM ENGR	HYDROGEN	61.7 MM	24.4%	0.200%	--	--	(1)
154)	MACHIMPLEX	CHINA (PRC)	HUMPHREYS & GLASGOW	METHANOL	22.9 MM	16.6%	11.500%	--	--	(1)
155)	ASIA OIL CO. LTD.	JAPAN	JAPAN GASOLINE	HYDROGEN	21.3 MM	23.1%	0.220%	--	--	(4)
156)	TOA OIL	JAPAN	CHIYODA CHEM ENGR	HYDROGEN	35.3 MM	17.2%	0.500%	--	--	(1)
157)	SIDERURG. VENEZOLANA	VENEZUELA	ARTHUR G MCKEE	RED. GAS	28.3 MM	20.7%	0.950%	--	--	(1)
158)	TOKUYAMA STYRENE	JAPAN	JAPAN GASOLINE	HYDROGEN	4.1 MM	11.5%	0.200%	--	--	(1)
159)	CHUNG TAI CHEMICAL	TAIWAN	DIDIER ENGRG GMBH	HYDROGEN	12.4 MM	20.2%	0.500%	--	--	(1)
160)	MONTEDISON S.P.A.	ITALY	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
161)	CHINA MAN-MADE FIBER	TAIWAN	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
162)	CNTIC NAXI	CHINA (PRC)	PULLMAN KELLOGG	AMMONIA	1000 TPD	17.7%	0.070%	--	--	(1)

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NOTES - (1) CONVENTIONAL BENFIELD (2) HIPURE (3) CATACARB CONVERSION (4) VETROCOKE CONVERSION (5) OTHER CONVERSION

BENFIELD CORP. GENERAL PLANT LIST

	PLANT OWNER	LOCATION	CONTRACTOR/ENGINEER	TYPE	*SIZE	CO2 IN	CO2 OUT	H2S IN	H2S OUT	NOTE
163)	CNTIC PANJIN	CHINA (PRC)	PULLMAN KELLOGG	AMMONIA	1000 TPD	17.7%	0.070%	--	--	(1)
164)	CNTIC WOLITUN	CHINA (PRC)	PULLMAN KELLOGG	AMMONIA	1000 TPD	17.7%	0.070%	--	--	(1)
165)	CNTIC TSANGCHOU	CHINA (PRC)	PULLMAN KELLOGG	AMMONIA	1000 TPD	17.7%	0.070%	--	--	(1)
166)	CNTIC SUICHIANG	CHINA (PRC)	PULLMAN KELLOGG	AMMONIA	1000 TPD	17.7%	0.070%	--	--	(1)
167)	CNTIC SISHUI	CHINA (PRC)	PULLMAN KELLOGG	AMMONIA	1000 TPD	17.7%	0.070%	--	--	(1)
168)	CNTIC ZHIJIANG	CHINA (PRC)	PULLMAN KELLOGG	AMMONIA	1000 TPD	17.7%	0.070%	--	--	(1)
169)	CNTIC YOUYANG	CHINA (PRC)	PULLMAN KELLOGG	AMMONIA	1000 TPD	17.7%	0.070%	--	--	(1)
170)	CNTIC CHENGDU	CHINA (PRC)	TOYO ENGINEERING	AMMONIA	1000 TPD	17.3%	0.100%	--	--	(1)
171)	CNTIC JINAN	CHINA (PRC)	TOYO ENGINEERING	AMMONIA	1000 TPD	17.3%	0.100%	--	--	(1)
172)	ZUARI AGRO	INDIA	PLANT OWNER	AMMONIA	725 TPD	22.3%	0.200%	--	--	(4)
173)	NATURAL GAS OF N.Z.	NEW ZEALAND	SCIEN. DESIGN LTD	NAT. GAS	15.6 MM	48.2%	2.000%	--	--	(1)
174)	SYNCRUDE CANADA LTD.	CANADA	BECHTEL CORP	HYDROGEN	90.0 MM	19.7%	0.150%	--	--	(1)
175)	SYNCRUDE CANADA LTD.	CANADA	BECHTEL CORP	HYDROGEN	90.0 MM	19.7%	0.150%	--	--	(1)
176)	SANTOS LTD TRAIN D	AUSTRALIA	PLANT OWNER	NAT. GAS	124.6 MM	16.0%	2.000%	0.005%	1 PPM	(1)
177)	CO SIGURIA GUANABARA	BRAZIL	FIRMA CARL STILL	RED. GAS	14.5 MM	21.5%	4.500%	0.010%	0 PPM	(1)
178)	CO SIGURIA GUANABARA	BRAZIL	FIRMA CARL STILL	RED. GAS	17.7 MM	3.9%	0.500%	0.580%	90 PPM	(1)
179)	FARMLAND INDUSTRIES	OKLAHOMA, USA	PULLMAN KELLOGG	AMMONIA	1250 TPD	18.3%	0.100%	--	--	(1)
180)	FARMLAND INDUSTRIES	KANSAS, USA	PLANT OWNER	AMMONIA	1250 TPD	18.3%	0.100%	--	--	(3)
181)	TAESUNG LUMBER	SOUTH KOREA	DAVY POWERGAS	METHANOL	5.6 MM	11.7%	1.500%	--	--	(1)
182)	PT PUPUK SRIWIDJAJA	INDONESIA	PULLMAN KELLOGG	AMMONIA	1000 TPD	18.4%	0.070%	--	--	(1)
183)	PT PUPUK SRIWIDJAJA	INDONESIA	PULLMAN KELLOGG	NAT. GAS	20.4 MM	18.2%	0.500%	--	--	(1)
184)	CONOCO KILLINGHOLME	GREAT BRITAIN	HUMPHREYS & GLASGOW	HYDROGEN	21.3 MM	14.1%	0.200%	--	--	(1)
185)	TECHMASH. TOGLIATTI	USSR	CHEMICO	AMMONIA	1500 TPD	18.0%	0.050%	--	--	(1)
186)	TECHMASH. TOGLIATTI	USSR	CHEMICO	AMMONIA	1500 TPD	18.0%	0.050%	--	--	(1)
187)	TECHMASH. TOGLIATTI	USSR	CHEMICO	AMMONIA	1500 TPD	18.0%	0.050%	--	--	(1)
188)	TECHMASH. TOGLIATTI	USSR	CHEMICO	AMMONIA	1500 TPD	18.0%	0.050%	--	--	(1)
189)	ASIA OIL COMPANY	JAPAN	DAVY POWERGAS	HYDROGEN	31.2 MM	23.6%	0.200%	--	--	(1)
190)	IRANIAN FERTILIZER	IRAN	HUMPHREYS & GLASGOW	AMMONIA	1320 TPD	18.2%	0.150%	--	--	(1)
191)	MARUZEN GAS DEVELOP.	JAPAN	PULLMAN KELLOGG	HYDROGEN	32.1 MM	24.3%	0.150%	--	--	(1)
192)	KYOWA PETROCHEMICAL	JAPAN	JAPAN GASOLINE	OXO GAS	10.5 MM	40.9%	0.180%	--	--	(1)
193)	NEW CENTRAL JUTE	INDIA	PLANT OWNER	SNG	60.0 MM	28.0%	0.800%	--	--	(1)
194)	COLUMBIA NYPRO	GEORGIA, USA	DAVY POWERGAS	HYDROGEN	11.3 MM	19.4%	0.100%	--	--	(1)
195)	CHEVRON SACROC II	TEXAS, USA	OLSEN ENGINEERING	NAT. GAS	35.0 MM	24.0%	1.000%	0.002%	1 PPM	(1)
196)	CNTIC	CHINA (PRC)	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
197)	PETROLEOS MEXICANOS	MEXICO	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
198)	ASHLAND PETROLEUM	KENTUCKY, USA	J F PRITCHARD CO	HYDROGEN	19.4 MM	21.0%	0.300%	--	--	(1)
199)	IMPERIAL CHEM INDUST	GREAT BRITAIN	HUMPHREYS & GLASGOW	AMMONIA	1250 TPD	17.8%	0.150%	--	--	(1)
200)	NIPPON PETROL. REF.	JAPAN	DAVY POWERGAS	HYDROGEN	39.5 MM	23.5%	0.180%	--	--	(1)
201)	INDUST ANGLAGEN INP	EAST GERMANY	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
202)	EGYPTIAN GEN'L ORGAN	EGYPT	FRIEDRICH UHDE	AMMONIA	1100 TPD	18.0%	0.100%	--	--	(1)
203)	TARRAGONA QUIMICA SA	SPAIN	FRIEDRICH UHDE	VIN. AC.	2.5 MM	10.0%	2.960%	--	--	(1)
204)	SIDOR I	VENEZUELA	PULLMAN SWINDELL	NAT. GAS	9.5 MM	11.3%	1.000%	0.003%	1 PPM	(1)
205)	BORCO DESULF. CO.	BAHAMAS	BADGER	HYDROGEN	25.0 MM	22.4%	0.500%	--	--	(1)
206)	FUJI OIL COMPANY	JAPAN	CHIYODA CHEM ENGR	HYDROGEN	41.2 MM	24.7%	0.300%	--	--	(1)
207)	SEMADCO/TALKHA II	EGYPT	FOST. WHEELER ITALY	AMMONIA	1250 TPD	17.3%	0.050%	--	--	(1)
208)	NATIONAL OIL CORP	LIBYA	FRIEDRICH UHDE	AMMONIA	1100 TPD	20.1%	0.100%	--	--	(1)

* - SIZE IS EITHER IN MILLIONS OF NORMAL CUBIC FEET OF PRODUCT GAS PER DAY OR IN TONS OF PRODUCT PER DAY

NOTES - (1) CONVENTIONAL BENFIELD (2) HIPURE (3) CATACARB CONVERSION (4) VETROCOKE CONVERSION (5) OTHER CONVERSION

	PLANT OWNER	LOCATION	CONTRACTOR/ENGINEER	TYPE	*SIZE	CO2 IN	CO2 OUT	H2S IN	H2S OUT	NOTE
209)	PERTAMINA - ARUN	INDONESIA	BECHTEL INT.	LNG	226.0 MM	14.7%	0.005%	0.008%	1 PPM	(2)
210)	PERTAMINA - ARUN	INDONESIA	BECHTEL INT.	LNG	226.0 MM	14.7%	0.005%	0.008%	1 PPM	(2)
211)	PERTAMINA - ARUN	INDONESIA	BECHTEL INT.	LNG	226.0 MM	14.7%	0.005%	0.008%	1 PPM	(2)
212)	ASIA KYOSEKI	JAPAN	CHIYODA CHEM ENGR	HYDROGEN	18.9 MM	22.8%	0.150%	--	--	(1)
213)	IRVING REFINING CO.	CANADA	FOSTER WHEELER CORP	HYDROGEN	39.8 MM	23.6%	0.150%	--	--	(1)
214)	MAHARASTRA COOP FERT	INDIA	HALDOR TOPSOE	AMMONIA	200 TPD	22.6%	0.100%	--	--	(1)
215)	NITRATOS DE CASTILLA	SPAIN	PLANT OWNER	AMMONIA	135 TPD	32.2%	0.060%	--	--	(4)
216)	NITRATOS DE CASTILLA	SPAIN	PLANT OWNER	AMMONIA	135 TPD	32.2%	0.060%	--	--	(4)
217)	PAKSAUDI FERTILIZERS	PAKISTAN	SNAM PROGETTI	AMMONIA	1100 TPD	18.7%	0.100%	--	--	(1)
218)	NITRIGIN EIRE TEOR.	IRELAND	KELLOGG INT.	AMMONIA	1500 TPD	17.5%	0.050%	--	--	(1)
219)	UK-WESSLING	WEST GERMANY	KRUPP KOPPERS	HYDROGEN	7.3 MM	9.1%	0.100%	--	--	(1)
220)	PT PUPUK SRIWIDJAJA	INDONESIA	PULLMAN KELLOGG	AMMONIA	1000 TPD	18.4%	0.070%	--	--	(1)
221)	PT PUPUK SRIWIDJAJA	INDONESIA	PULLMAN KELLOGG	NAT. GAS	20.4 MM	18.2%	0.500%	--	--	(1)
222)	FARMLAND INDUSTRIES	OKLAHOMA, USA	PULLMAN KELLOGG	AMMONIA	1250 TPD	18.3%	0.100%	--	--	(3)
223)	FARMLAND INDUSTRIES	LOUISIANA, USA	PULLMAN KELLOGG	AMMONIA	1250 TPD	18.3%	0.100%	--	--	(3)
224)	PT PUPUK KUJANG	INDONESIA	PULLMAN KELLOGG	AMMONIA	1100 TPD	19.0%	0.100%	--	--	(1)
225)	COAL MIN'G RES. CENT.	JAPAN	HITACHI LIMITED	COAL GAS	0.5 MM	7.4%	2.000%	0.110%	90 PPM	(1)
226)	LIZENSINTORG MOSKOV	USSR	FRIEDRICH UHDE	VIN. AC.	140.0 MM	20.2%	2.500%	--	--	(1)
227)	NYPRO (U.K.) LTD	GREAT BRITAIN	KINETICS TECH. INT.	HYDROGEN	14.9 MM	18.9%	0.100%	--	--	(1)
228)	COLUMBIA NITROGEN	GEORGIA, USA	C.F. BRAUN	AMMONIA	1500 TPD	17.1%	0.080%	--	--	(1)
229)	CHINA PETROLEUM CORP	TAIWAN	ABR ENGINEERING	NAT. GAS	44.5 MM	43.6%	5.000%	0.002%	0 PPM	(1)
230)	SIDOR II	VENEZUELA	PULLMAN SWINDELL	NAT. GAS	52.0 MM	11.3%	1.000%	--	--	(1)
231)	TOKYO GAS NEGISHI	JAPAN	DAVY POWERGAS	SNG	38.7 MM	21.1%	1.000%	--	--	(1)
232)	TOKYO GAS NEGISHI	JAPAN	DAVY POWERGAS	SNG	38.7 MM	21.1%	1.000%	--	--	(1)
233)	TECHNOIMPORT	BULGARIA	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
234)	PETRO. RIO TERCERO	ARGENTINA	SOC. TECH. ENT. CHIM.	HYDROGEN	4.8 MM	8.0%	0.100%	--	--	(1)
235)	STATE FERT MFG CO	SRI LANKA	PULLMAN KELLOGG	AMMONIA	650 TPD	19.0%	0.100%	--	--	(1)
236)	HINDUSTAN STEEL LTD	INDIA	FERT. CORP. OF INDIA	AMMONIA	180 TPD	23.0%	0.100%	--	--	(1)
237)	INDIAN FARM FERT CO	INDIA	PULLMAN KELLOGG	AMMONIA	1000 TPD	22.8%	0.050%	--	--	(1)
238)	NEYVELI LIGNITE CORP	INDIA	FERT. CORP. OF INDIA	AMMONIA	350 TPD	6.5%	0.005%	1.100%	0 PPM	(2)
239)	NEYVELI LIGNITE CORP	INDIA	FERT. CORP. OF INDIA	AMMONIA	350 TPD	32.0%	0.050%	--	--	(1)
240)	TECHMASH ROSSOSH	USSR	TOYO ENGINEERING	AMMONIA	1500 TPD	17.4%	0.050%	--	--	(1)
241)	TECHMA NOVOMOSKOVSK	USSR	TOYO ENGINEERING	AMMONIA	1500 TPD	17.4%	0.050%	--	--	(1)
242)	TECHMASH GRODNO	USSR	TOYO ENGINEERING	AMMONIA	1500 TPD	17.4%	0.050%	--	--	(1)
243)	TECHMASH CHERKASSY	USSR	TOYO ENGINEERING	AMMONIA	1500 TPD	17.4%	0.050%	--	--	(1)
244)	SO AFRI COAL OIL GAS	SOUTH AFRICA	LINDE A.G.	COAL GAS	262.0 MM	11.5%	0.008%	--	--	(2)
245)	SO AFRI COAL OIL GAS	SOUTH AFRICA	LINDE A.G.	COAL GAS	262.0 MM	11.5%	0.008%	--	--	(2)
246)	QATAR FERTILIZER CO	QATAR	DAVY POWERGAS	AMMONIA	1000 TPD	18.8%	0.200%	--	--	(1)
247)	KUWAIT OIL COMPANY	KUWAIT	KELLOGG INT.	NAT. GAS	61.5 MM	7.3%	0.400%	8.020%	700 PPM	(1)
248)	GULF OIL CANADA	CANADA	FOSTER WHEELER CORP	HYDROGEN	14.6 MM	15.9%	0.100%	--	--	(1)
249)	ASHUGANJ FERT & CHEM	BANGLADESH	FRIEDRICH UHDE	AMMONIA	1200 TPD	17.7%	0.100%	--	--	(1)
250)	COLORADO INTERSTATE	WYOMING, USA	STEARNS ROGER INC	NAT. GAS	44.6 MM	17.6%	0.025%	3.760%	3 PPM	(2)
251)	AZOT SANAYITAS	TURKEY	KELLOGG INT.	AMMONIA	1100 TPD	22.4%	0.050%	--	--	(1)
252)	NIPPON SHOKUBAI	JAPAN	HUMPHREYS & GLASGOW	ETH. OX.	35.5 MM	8.1%	1.300%	--	--	(1)
253)	OXITENO NORDESTE	BRAZIL	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)
254)	PPG INDUSTRIES	TEXAS, USA	SCIENTIFIC DESIGN	ETH. OX.	OPERATING DETAILS ARE NOT TO BE DISCLOSED					(1)

* - SIZE IS EITHER IN MILLIONS OF NORMAL CUBIC FEET OF PRODUCT GAS PER DAY OR IN TONS OF PRODUCT PER DAY

NOTES - (1) CONVENTIONAL BENFIELD (2) HIPURE (3) CATACARB CONVERSION (4) VETROCOKE CONVERSION (5) OTHER CONVERSION

BENFIELD CORP.

GENERAL PLANT LIST

	PLANT OWNER	LOCATION	CONTRACTOR/ENGINEER	TYPE	*SIZE	CO2 IN	CO2 OUT	H2S IN	H2S OUT	NOTE
255)	JEFFERSON CHEMICAL	TEXAS, USA	SCIENTIFIC DESIGN	ETH. OX	OPERATING DETAILS ARE CONFIDENTIAL					(1)
256)	FCI TROMBAY V	INDIA	SNAM PROGETTI	AMMONIA	1100 TPD	21.9%	0.100%	—	—	(1)
257)	DZERZHINSK ETH. OX	USSR	SCIENTIFIC DESIGN	ETH. OX	OPERATING DETAILS ARE CONFIDENTIAL					(1)
258)	POLIMEX	POLAND	HALDOR TOPSOE	AMMONIA	825 TPD	18.8%	0.100%	—	—	(1)
259)	POLIMEX	POLAND	HALDOR TOPSOE	AMMONIA	825 TPD	18.8%	0.100%	—	—	(1)
260)	PETROMIN	SAUDI ARABIA	TIELROOY & ASSOC	HYDROGEN	61.5 MM	15.9%	0.100%	—	—	(1)
261)	TECHMASHIMPORT PERM	USSR	TOYO ENGINEERING	AMMONIA	1500 TPD	17.7%	0.050%	—	—	(1)
262)	MINISTRY CHEM INDUST	ROMANIA	SCIENTIFIC DESIGN	ETH. OX	OPERATING DETAILS ARE CONFIDENTIAL					(1)
263)	I.N A NAFTAPLAN	YUGOSLAVIA	KINETICS TECH INT	NAT. GAS	27.2 MM	24.5%	3.600%	—	—	(1)
264)	PHILADELPHIA GAS WKS	PENNA., USA	WOODALL DUCKHAM	SNG	55.3 MM	24.5%	1.600%	—	—	(1)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	4.5 MM	24.1%	0.500%	—	—	
265)	DOW CHEMICAL CANADA	CANADA	DOW CANADA	ETH. OX	OPERATING DETAILS ARE CONFIDENTIAL					(1)
266)	KEIYO GAS COMPANY	JAPAN	JAPAN GASOLINE	SNG	16.9 MM	20.7%	2.000%	—	—	(1)
267)	KASEI MIZUSHIMA	JAPAN	JAPAN GASOLINE	OXO GAS	12.9 MM	30.6%	0.800%	—	—	(1)
268)	I.N.A NAFTAPLAN	YUGOSLAVIA	KELLOGG CONT. BV.	AMMONIA	1500 TPD	19.8%	0.050%	—	—	(1)
269)	TECHMASHIMP. ANGARSK	USSR	TOYO ENGINEERING	AMMONIA	1500 TPD	21.9%	0.050%	—	—	(1)
270)	SANTOS LTD TRAIN E	AUSTRALIA	PLANT OWNER	NAT. GAS	130.8 MM	16.0%	2.000%	—	—	(1)
271)	JEFFERSON CHEMICAL	TEXAS, USA	SCIENTIFIC DESIGN	ETH. OX	OPERATING DETAILS ARE CONFIDENTIAL					(1)
272)	JEFFERSON CHEMICAL	TEXAS, USA	SCIENTIFIC DESIGN	ETH. OX	OPERATING DETAILS ARE CONFIDENTIAL					(1)
273)	AGRICO CHEMICAL	ARKANSAS, USA	PLANT OWNER	AMMONIA	1000 TPD	18.2%	0.080%	—	—	(3)
274)	PETROLEOS MEXICANOS	MEXICO	SCIENTIFIC DESIGN	ETH. OX	OPERATING DETAILS ARE CONFIDENTIAL					(1)
275)	BALTIMORE GAS & ELEC	MARYLAND, USA	PLANT OWNER	SNG	56.1 MM	23.4%	0.500%	—	—	(5)
	LINE ABOVE IS MAIN STREAM. THIS LINE FOR SECOND ABSORBER			HYDROGEN	4.0 MM	23.6%	1.000%	—	—	
276)	UNION OIL CALIFORNIA	ALABAMA, USA	ORTLOFF ENGINEERING	NAT. GAS	51.1 MM	4.6%	0.020%	0.920%	1 PPM	(2)
277)	GENERAL ELECTRIC	NEW YORK, USA	C. F. BRAUN	CONFID.	OPERATING DETAILS ARE CONFIDENTIAL					(1)
278)	NATIONAL OIL LIBYA	LIBYA	SNAM PROGETTI	AMMONIA	1100 TPD	19.9%	0.100%	—	—	(1)
279)	NIZHNEKAMSK ETH. OX	USSR	SCIENTIFIC DESIGN	ETH. OX	OPERATING DETAILS ARE CONFIDENTIAL					(1)
280)	FAUJI FERTILIZER	PAKISTAN	SNAM PROGETTI	AMMONIA	1100 TPD	18.8%	0.100%	--	--	(1)
281)	TOKYO GAS	JAPAN	HITACHI LIMITED	SNG	1.2 MM	21.5%	0.090%	--	--	(1)
282)	PAKISTAN AJMAN FERT	PAKISTAN	HALDOR TOPSOE	AMMONIA	1100 TPD	17.5%	0.100%	--	--	(1)
283)	I E L - KANPUR	INDIA	PLANT OWNER	AMMONIA	457 TPD	22.7%	0.200%	--	--	(1)
284)	FAZINKA - FAZIN	YUGOSLAVIA	SCIENTIFIC DESIGN	ETH. OX	OPERATING DETAILS ARE CONFIDENTIAL					(1)
285)	PLACID OIL	LOUISIANA, USA	B & C ASSOCIATES	NAT. GAS	56.7 MM	7.3%	1.400%	--	--	(1)
286)	KALTIMA AMMONIA	INDONESIA	LURGI KOHLE UND MIN.	AMMONIA	1500 TPD	19.1%	0.130%	--	--	(1)
287)	HOKORIKU GAS	JAPAN	DAVY POWERGAS	TOWN GAS	13.9 MM	18.6%	2.000%	--	--	(1)
288)	HINDUSTAN FERTILIZER	INDIA	FERT. CORP. OF INDIA	AMMONIA	600 TPD	18.6%	0.100%	--	--	(1)
289)	SO AFRI COAL OIL GAS	SOUTH AFRICA	LINDE A.G.	COAL GAS	262.6 MM	11.5%	0.008%	--	--	(2)
290)	SO AFRI COAL OIL GAS	SOUTH AFRICA	LINDE A G	COAL GAS	262.6 MM	11.5%	0.008%	--	--	(2)
291)	ODUR FADRIKA SECERA	YUGOSLAVIA	SIRYCON LIMITED	CITR AC.	4.4 MM	5.0%	0.500%	--	--	(1)
292)	POLYCHEM	INDIA	FRIEDRICH UHDE	VIN AC.	OPERATING DETAILS ARE CONFIDENTIAL					(1)
293)	PETROCHEMICAL INDUST	KUWAIT	HALDOR TOPSOE	AMMONIA	1100 TPD	19.4%	0.100%	--	--	(1)
294)	KYODO OXYGEN	JAPAN	MARUZEN ENGINEERING	CO2 REC.	6.5 MM	46.0%	0.600%	--	--	(1)
295)	STATE ORGAN OIL PROJ	IRAQ	KINETICS TECH. INT.	HYDROGEN	65.4 MM	23.2%	0.200%	--	--	(1)
296)	IDEMITSU PETROCHEM	JAPAN	TOYO ENGINEERING	HYDROGEN	6.3 MM	3.9%	0.100%	--	--	(1)
297)	AZZAHIYA OIL RFG CO	LIBYA	SCIENTIFIC DESIGN	ETH OX	OPERATING DETAILS ARE CONFIDENTIAL					(1)
298)	BALEH PAPAN	INDONESIA	UNIVERSAL OIL PROD	HYDROGEN	33.8 MM	21.7%	0.100%	--	--	(1)
299)	BALEH PAPAN	INDONESIA	UNIVERSAL OIL PROD	HYDROGEN	33.8 MM	21.7%	0.100%	—	—	(1)

* - SIZE IS EITHER IN MILLIONS OF NORMAL CUBIC FEET OF PRODUCT GAS PER DAY OR IN TONS OF PRODUCT PER DAY

NOTES - (1) CONVENTIONAL BENFIELD (2) HIPURE (3) CATACARB CONVERSION (4) VETROCOKE CONVERSION (5) OTHER CONVERSION

	PLANT OWNER	LOCATION	CONTRACTOR/ENGINEER	TYPE	*SIZE	CO2 IN	CO2 OUT	H2S IN	H2S OUT	NOTE
300)	DUMAI REFINERY	INDONESIA	UNIVERSAL OIL PROD	HYDROGEN	30.9 MM	10.8%	0.100%	--	--	(1)
301)	DUMAI REFINERY	INDONESIA	UNIVERSAL OIL PROD	HYDROGEN	30.9 MM	10.8%	0.100%	--	--	(1)
302)	POWERINE OIL CO	CALIFORNIA, USA	FLUOR CORPORATION	HYDROGEN	10.3 MM	20.6%	0.200%	--	--	(1)
303)	ABU DHABI NAT'L OIL	ABU DHABI	HALDOR TOPSOE	AMMONIA	1100 TPD	10.3%	0.100%	--	--	(1)
304)	NIPPON PET RFG CORP	JAPAN	CHIYODA CHEM ENGR	HYDROGEN	40.5 MM	23.5%	0.200%	--	--	(1)
305)	KUWAIT NAT. PET. CO.	KUWAIT	FOSTER WHEELER CORP	HYDROGEN	40.5 MM	19.9%	0.120%	--	--	(1)
306)	KUWAIT NAT. PET. CO.	KUWAIT	FOSTER WHEELER CORP	HYDROGEN	40.5 MM	19.9%	0.120%	--	--	(1)

* - SIZE IS EITHER IN MILLIONS OF NORMAL CUBIC FEET OF PRODUCT GAS PER DAY OR IN TONS OF PRODUCT PER DAY

NOTES - (1) CONVENTIONAL BENFIELD (2) HIPURE (3) CATACARB CONVERSION (4) VETROCOKE CONVERSION (5) OTHER CONVERSION

2/81

STATUS SUMMARY

CNG ACID GAS REMOVAL PROCESS

1.0 GENERAL INFORMATION

Developer: Consolidated Natural Gas Research Company
Cleveland, OH 44106

Licensor: Consolidated Natural Gas Research Company
Cleveland, OH 44106

Type: The CNG process is a low temperature physical absorption process that removes acid gases (CO_2, H_2S, COS) and other trace contaminants from medium to high pressure gas streams containing substantial amounts of CO_2. The process has three novel features:

(i) Absorption of sulfur-compounds and trace contaminants with pure liquid CO_2.

(ii) Regeneration of pure CO_2 with simultaneous concentration of H_2S and trace contaminants by triple-point crystallization.

(iii) Absorption of CO_2 with a slurry of organic liquid containing solid CO_2.

Conditions: The sulfur absorber operates at -67°F, slightly above the triple point. The feed gas pressure can range from 300 to 1500 psia. However, the CO_2 dew point of the feed gas must be above the triple point temperature. The feed gas must have a CO_2 partial pressure of at least 75 psia. The triple-point crystallizer operates at pressures and temperatures near the triple point of CO_2. The CO_2 absorber operates at temperature ranges between -67°F and -112°F.

Applications: The CNG process can be applied to process gas containing medium to high CO_2 content, e.g., synthesis gas produced by coal gasification, heavy oil partial oxidation, and natural gas in enhanced oil recovery projects.

Status: The process is currently at the bench scale stage of development. Future plans call for a 2-3 ton/day Process Development Unit (PDU).

2.0 PROCESS DEVELOPMENT

- Development work on the CNG acid gas removal process has been sponsored since 1973 by the CNG Research Company

- A U.S. Patent 4,270,937 was filed in 1979 and was accepted in June 1981.

- Since 1981, the CNG process development work was co-sponsored by the U. S. Department of Energy, and CNG Research Company.

- To date, an extensive process data base has been accumulated which includes:

  (i) Vapor-liquid equilibrium data for binary systems of CO_2 with sulfur-compounds and with ethers and ketones, and multi-component systems including hydrogen, carbon monoxide, methane, carbon dioxide, and hydrogen sulfide.

  (ii) A bench-scale slurry formation and melting apparatus which provides data on slurry pumping and on rates of slurry formation and melting.

  (iii) A bench-scale closed-cycle triple point crystallizer which provides data on rates of crystallization and melting and on separation factors attainable in a single stage of crystallization.

3.0 SOLVENT CHARACTERISTICS

- Physical and Chemical Properties of CO_2

Triple Point	-69.9°F, 75.1 psia
Sublimation Temperature	109.3°F
Specific gravity liquid @ 67°F	1.18
Viscosity @ -67oF	0.35
Surface Tension @ -67°F	10 dynes/cm
Molecular Weight	44
Heat of Fusion	86 Btu/lb
Heat of Vaporization	150 Btu/lb

(Source Ref. 1 & 2)

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4.0 PROCESS DESCRIPTION

The CNG process consists of five main steps. (Figure 4-1).

- Precooling, Water Removal

  The raw gas is cooled and residual water vapor is removed in a dehydration system to prevent subsequent icing. The water-free crude gas is further cooled to its carbon dioxide dew point (-56°C) by countercurrent heat exchange with return clean gas and carbon dioxide.

- Carbon Dioxide Condensation, Sulfur-Compound Absorption

  Carbon dioxide is partially condensed by cooling the gas from its dew point to about -55°C. Depending on CO_2 dew point, a fraction of the CO_2 in the crude gas stream is condensed together with the sulfur-compounds. The gas at -55°C is scrubbed by liquid CO_2 to remove H_2S, COS and other trace impurities from the feed gas. The absorption is essentially isothermal since the heat of absorption is being dissipated as heat of vaporization of a small portion of the liquid CO_2.

  The liquid CO_2 with all the sulfur-compounds, other trace contaminants and some co-absorbed light hydrocarbons is combined with the contaminated liquid CO_2 condensed in precooling the raw gas. The light hydrocarbons are stripped from this combined liquid CO_2 stream and recycled and mixed with the feed gas. Any higher hydrocarbons (C_4-C_6) in the feed gas will remain with the condensed CO_2.

- Carbon Dioxide Regeneration by Triple Point Crystallization.

  The contaminated liquid CO_2 stream leaving the light ends stripping tower is processed in a direct-contact triple point crystallizer with vapor compression. Solid CO_2 is formed by adiabatic flashing of the liquid CO_2 stream near the top of the crystallizer. An H_2S-rich gas stream is produced and is continuously withdrawn from the top. All C_2-C_6 hydrocarbons to the crystallizer are removed with the H_2S rich stream. The solid CO_2 crystals fall to the bottom of the crystallizer where they are melted by direct contact with condensing carbon dioxide vapor. Pure carbon dioxide liquid thus produced is split into two streams, one in absorbent recycled to the H_2S absorber, the other is sent back through the process for refrigeration and power recovery, and is delivered as a product stream or vented to the atmosphere.

- Carbon Dioxide Removal

  CO_2 remaining in the gas after removal of sulfur-compounds is absorbed at temperatures below the CO_2 triple-point with a slurry absorbent. The slurry absorbent is a saturated solution of an organic solvent and CO_2 containing suspended particles of solid CO_2. As CO_2 is absorbed (condensed), the latent heat released melts the solid CO_2 contained in the slurry absorbent. The direct refrigeration provided by the melting of solid CO_2 enables a small absorbent flow to accommodate the considerable heat of condensation and absorption of the CO_2 vapor. The cold, purified gas stream then leaves the acid gas removal process after heat exchange with the raw gas stream.

- Regeneration of Carbon Dioxide Absorbent Slurry

  The CO_2-rich solvent leaving the CO_2 absorber near the triple-point temperature contains no solid CO_2. This stream is flashed in a drum to vaporize any methane or other light components. The CO_2-rich absorbent is next cooled by external refrigeration and then flashed to lower pressure in a number of stages, so as to generate a cold slurry of liquid solvent and solid CO_2. Nitrogen stripping of the solvent may sometimes be required to produce a very lean solvent. The regenerated slurry absorbent is recirculated to the CO_2 absorber while the CO_2 flashed gas is vented to the atmosphere after recovery of refrigeration and power.

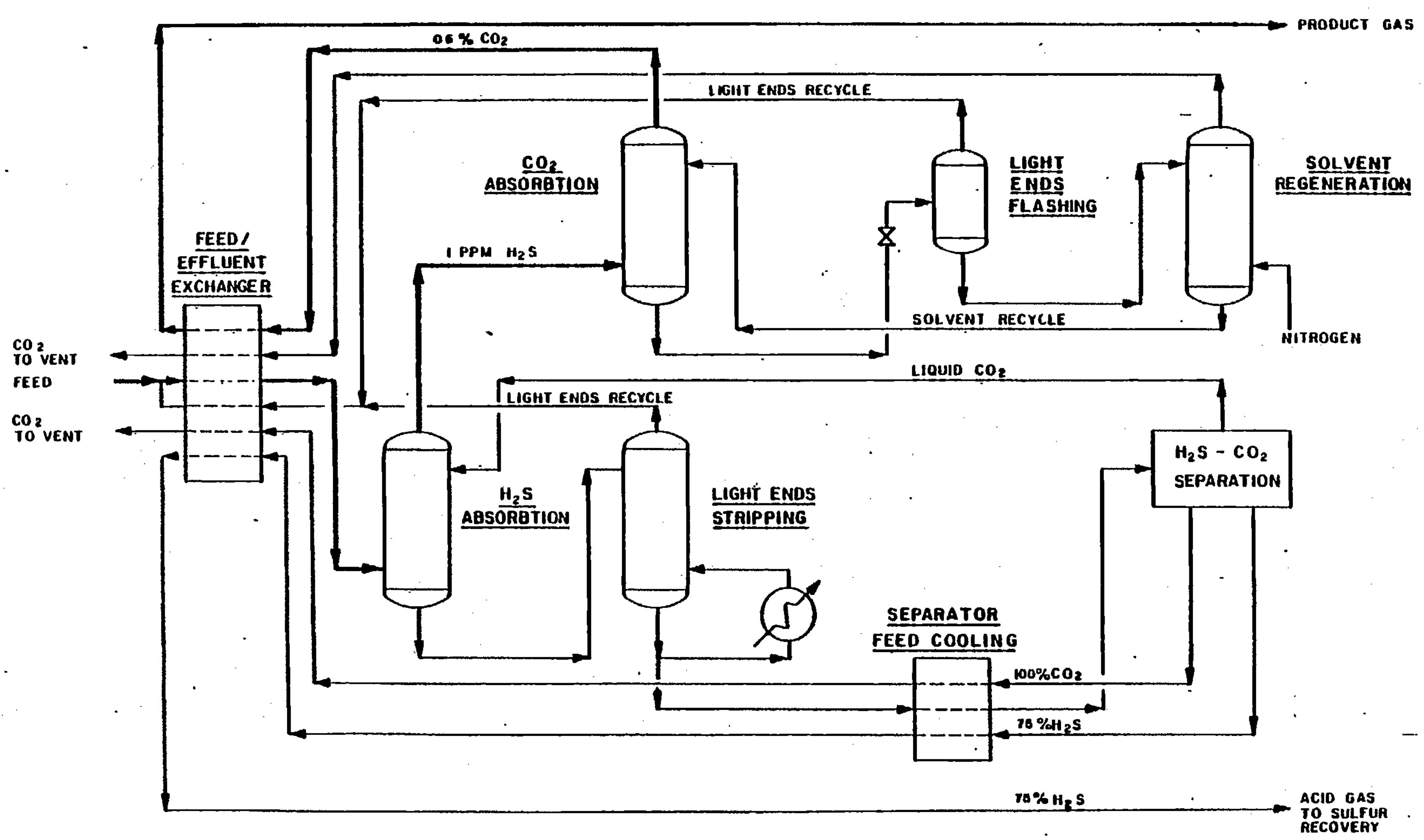

FIGURE 4-1
SIMPLIFIED PROCESS FLOW DIAGRAM
(Source: Ref. 3)

5.0 COMMERCIAL DESIGN DATA

A summary of the design data for a CNG acid gas removal system for a 250 Billion Coal-to-SNG plant is shown in Table 5-1. The feed gas is from the gasification of a U. S. Eastern Coal in the BI-GAS gasifiers. Table 5-2 gives the calculation of contribution to gas cost for the case presented in Table 5-1.

6.0 PROCESS CHARACTERISTICS

- Advantages
  - Pure liquid CO_2 absorbs COS better than H_2S. An absorber for sulfur-compounds designed to remove H_2S using liquid CO_2 will also remove COS and all other less volatile sulfur-compounds in the gas.
  - A sulfur-rich stream up to 75% H_2S can be produced by triple point crystallization with low energy consumption.
  - A pure CO_2 stream with 1 ppm impurities can be produced by triple point crystallization.
  - Low flowrate of liquid CO_2 used to absorb sulfur-compounds and other trace impurities due to: 1) unique COS absorption capability of liquid CO_2, 2) low absorption temperature and 3) low molecular weight of CO_2 which favors high gas solubility per unit volume of solvent.
  - Low flowrate of slurry absorbent for CO_2 removal due to: 1) low temperature of absorption, 2) prior bulk condensation of CO_2, and 3) high effective heat capacity of the slurry containing solid CO_2 (as in situ phase change refrigerant).
  - The CNG process utilizes relatively small temperature and pressure driving forces to effect the desired separations. Therefore the CNG process has a low energy consumption.
  - The CNG process uses liquid CO_2, a byproduct of the process, as the absorbent for sulfur-compounds removal. This can reduce the solvent makeup requirements.
- Disadvantages
  - Separate raw gas dehydration and condensible hydrocarbon cleanup systems are required prior to the CNG process.
  - Low operating temperatures require extensive use of 3-1/2% Nickel Steel.

TABLE 5-1

DESIGN DATA FOR A CNG AGR SYSTEM IN A 250 BILLION BTU/DAY COAL-TO-SNG PLANT

Stream	Shifted Syngas	Acid gas to Claus	Stripper Overhead	CO2 Product	Sweet Gas	N2 to Stripper
Flowrate, Moles/hr	101,176	1,745	4,526	30,076	67,546	2,800
Composition (Vol %)						
CO	11.85	0.00	0.17	0.00	17.74	
H2	40.63	0.00	44 ppm	0.00	60.85	
CH4	13.94	0.00	6.43	0.00	20.45	
C2H6						
N2	0.22	0.00	61.36	0.00	0.36	100.0
CO2	31.99	24.99	32.04	100.00	0.60	
COS	0.02	1.35	0.00	0.00	< 1 ppm	
H2S	1.27	73.66	0.00	0.00	< 1 ppm	
H2O	0.08	0.00	0.00	0.00	0.00	

UTILITY REQUIREMENTS

Electric Power	14,651 Kw (Pumps & Comp) 2,507 Kw (Refrigeration)
Make-up Solvent	1.5 MM $/yr

CAPITAL COSTS

Equipment (Installed)	$ 90.6 MM (Mid-1982)
Royalties	$ 1.4 MM (Mid-1982)

TABLE 5-2

CALCULATION OF CONTRIBUTION TO GAS COST
CNG (SELECTIVE) CASE

Coal Type	Eastern U.S. (Bigas Gasifiers)
Evaluator	C.F.Braun
Project Report No.	GRI-81/0174
Date Published	May,1983
Plant Capacity	250 Billion Btu/day SNG

CAPITAL COSTS :	$ MM (Mid-1982)
Installed Equipment	90.6
Contingency @ 15%	13.6
Direct Facility Constr Investment	104.2
Home-Office costs @ 12%	12.5
Total Facility Constr Investment	116.7
Royalties	1.4
Initial Solvent Charge	8.4
Total Plant Investment	126.5

OPERATING COSTS :		$/hr
Electricity	17,250 Kw @ $0.05/Kwh	862.5
Solvent Make-up		173.5
TOTAL		1036.0

Total Operating Cost, $ MM/yr at 100 % Stream factor = 9.08 MM $/Yr

CONTRIBUTION TO GAS COSTS :	Specific Cost, $/MM Btu-Yr	Charge Rate, Year	Contribution, $/MM Btu
Capital Related	1.54	0.089	0.14
Operating	0.11	1.000	0.11
Total			0.25

- The process is not suitable for gases having low CO_2 concentration.

- Slurry systems may be difficult to operate.

o Hydrocarbons Losses: (Ref. 3)

Component	Losses as Percent of Feed
CO	<0.1%
H_2	Negligible
CH_4	2%

7.0 REFERENCES

1. Hise, R. E., L. G. Massey, R. J. Adler, C. B. Brosilow, N. C. Gardner, W. R. Brown, W. J. Cook, and M. Petrik, "The CNG Process: A New Approach to Physical Absorption Acid Gas Removal", AIChE Symposium Series No. 224, Vol. 79, 1983.

2. Cook W. J., W. R. Brown, Y.C. Lin, and M. Petrik", Refrigerant Absorbents for use in Acid Gas Removal", Paper presented at the AIChE 1983 Summer National Meeting, August 1983, Denver, Colorado.

3. "Fossil Fuel Gasification Technical Evaluation Services" - Evaluation of CNG Acid Gas Removal Process" Final Report, Prepared for the Gas Research Institute by CF Braun & Co., March 1983.

STATUS SUMMARY

RYAN-HOLMES ACID GAS REMOVAL PROCESSES

1.0 GENERAL INFORMATION

Developer: Koch Process Systems Inc.
Westborough, Massachusetts

Licensor: Koch Process Systems, Inc.
Westborough, Massachusetts

Type: The Ryan-Holmes Processes are low temperature distillation processes used for separating acid gases (CO_2 and sulfur compounds) from natural gas streams. An NGL-based additive, typically a C_4+ mixture, is introduced to or near the top of the distillation column to significantly alter the phase equilibrium of the systems in order to provide advantages in the distillations.

Conditions: The distillation columns operate between 350 to 650 psia. Depending on the scheme, the operating temperatures can range from -120°F to 40°F. Acid gas contents (mainly CO_2) in the feed can range from 30% to 90%.

Applications: The Ryan-Holmes processes can be applied to a variety of gas processing applications:

- Processing of associated gases from enhanced oil recovery projects;
- NGL recovery fom sour hydrocarbon gases;
- Upgrading BTU content and quality of natural gas stream with high acid gas concentration;
- Feed purification prior to follow-on cryogenic processing;

Process Schemes: Several different process schemes are possible with combinations of a demethanizer, an ethane recovery column, an NGL (additive) recovery column, and a CO_2 recovery column. (Refer to Section 4.0 for details)

Status: Koch Process Systems has licensed four applications of the Ryan/Holmes technology in Enhanced Oil Recovey projects. This process has not yet been applied, or proven to process synthesis gases from coal gasification units.

2.0 PROCESS DEVELOPMENT

- In 1978, a laboratory program consisting of multi-component phase equilibrium cell experiments was initiated.

- Vapor-liquid equilibrium data were studied for:

  - Methane-CO_2 at high CO_2 concentrations and near the freezing point of CO_2;
  - CO_2-C_2H_6-C_4 liquids and CO_2-H_2S-C_4 liquids over wide composition ranges.

- Experimental data were fitted using the Soave-Redlich-Kwong and Peng-Robinson equations-of-state.

- In 1981, a 600,000 SCFD pilot plant was installed at the ARCO Oil and Gas Company gas plant in Covington, Oklahoma. A variety of feed compositions were tested, including variation of CO_2 content from 10% to 90%, variation in NGL content and the presence of H_2S in feed.

- After completion of the pilot plant program, the pilot plant was relocated to Koch Refining Company's refinery in Corpus Christi, Texas.

- Further testing is planned to include effects of column internals on hydraulic characteristics and applications in the production of SNG and synthesis gas.

3.0 SOLVENT CHARACTERISTICS

The additives used in the Ryan Holmes processes are usually Natural Gas Liquids (NGL) extracted from natural gas feed. Typical additives most commonly used are C_4^+ mixtures. The use of NGL has the advantage that it is readily available and that it does not introduce any foreign components into the system. The functions of the additives in different separations are as follow:

- Separation of methane and CO_2 -
  - (i) prevents CO_2 freezing in the condenser;
  - (ii) raises the operating temperature of the distillation column, hence reducing refrigeration horsepower requirements;
  - (iii) increases the methane/CO_2 relative volatility above that for the binary system;
  - (iv) permits higher pressure operation;
- Separation of CO_2 from ethane-plus hydrocarbons -
  - (i) breaks the CO_2-ethane azeotrope;
  - (ii) allows higher operating temperatures;
- Separation of CO_2 and H_2S -
  - (i) enhances the CO_2/H_2S relative volatility.

Depending on the feed gas compositions and the type of application, the molar ratio of additive to methane content in feed gas can range from 0.25 to 4.5.

4.0 PROCESS DESCRIPTION

Depending on the application, several different Ryan-Holmes process flow schemes are possible:

- Enhanced Oil Recovery Gas Processing Facilities

  This system consists of three columns: demethanizer, ethane recovery column, and NGL product column, as shown in Figure 4-1.

  Dry feed is cooled by appropriate cold products and refrigeration before entering the demethanizer. An optimum NGL mixture is added in the demethanizer condenser, which operates at about -120°F. Product sales gas (H_2S-less than 4 ppm, CO_2-100 ppm to 2%) is produced overhead at essentially column pressure (normally 450-650 psi).

  The bottoms is fed to the ethane recovery column after heat exchange. This column normally operates at 350 to 550 psig with an overhead temperature of about 5°F. The additive NGL is introduced to the column several trays down from the top. CO_2 is produced with 50 ppm or less H_2S.

  The bottoms product from the ethane column is fed to the NGL recovery column. A light hydrocarbon is produced overhead which contains all the H_2S and CO_2 fed to the column. The bottom stream from this column is used as the distillation additive for the first two columns. A slip stream is taken from the bottom stream to remove the net C_4^+ contained in the feed.

  An alternative process scheme using four columns can also be used in this application. The four-column arrangement has lower energy requirements and the overall capital costs are also lower. This system, as shown in Figure 4-2, consists of a demethanizer, an ethane recovery column, a CO_2 recovery column, and an NGL (additive) recovery column.

  Dry feed enters the ethane recovery column where the split betweeen CO_2 overhead product and a bottoms product stream consisting of C_2, H_2S, C_3^+ and additive, is effected. The overhead is then compressed, chilled by refrigerant and fed to the CO_2 recovery column. No additive is used in the CO_2 recovery column so the overhead is a methane stream with 15 to 30% CO_2. The bottoms product is pure liquid CO_2 and can be pumped to the required reinjection pressure. The overhead from the CO_2 recovery column next flows to the demethanizer where additive is used to produce a sales gas quality fraction overhead. Less additive is required for the four column scheme since much of the CO_2 has been removed in the CO_2 recovery column.

The bottoms of the demethanizer contains CO_2 plus additive. It is used to fulfill some of the additive requirement in the ethane recovery column with the remainder of that requirement being made up with fresh recycle additive.

The bottoms from the ethane recovery column flows to the additive recovery column in which a light NGL-heavy NGL split is made.

- Upgrading BTU Content and Quality of Sweet Natural Gas Streams

  Only two columns, a demethanizer and an additive recovery column, are required; see Figure 4-3. The second column operates as a conventional distillation column.

- Recovery of NGL without Methane/CO2 Separation

  This process scheme applies where combined methane-CO_2 overhead can either be reinjected or burned directly as a low BTU fuel. Ethane and C_3 hydrocarbons are recovered in the bottoms of the demethanizer and processed in the additive recovery column, as shown in Figure 4-4.

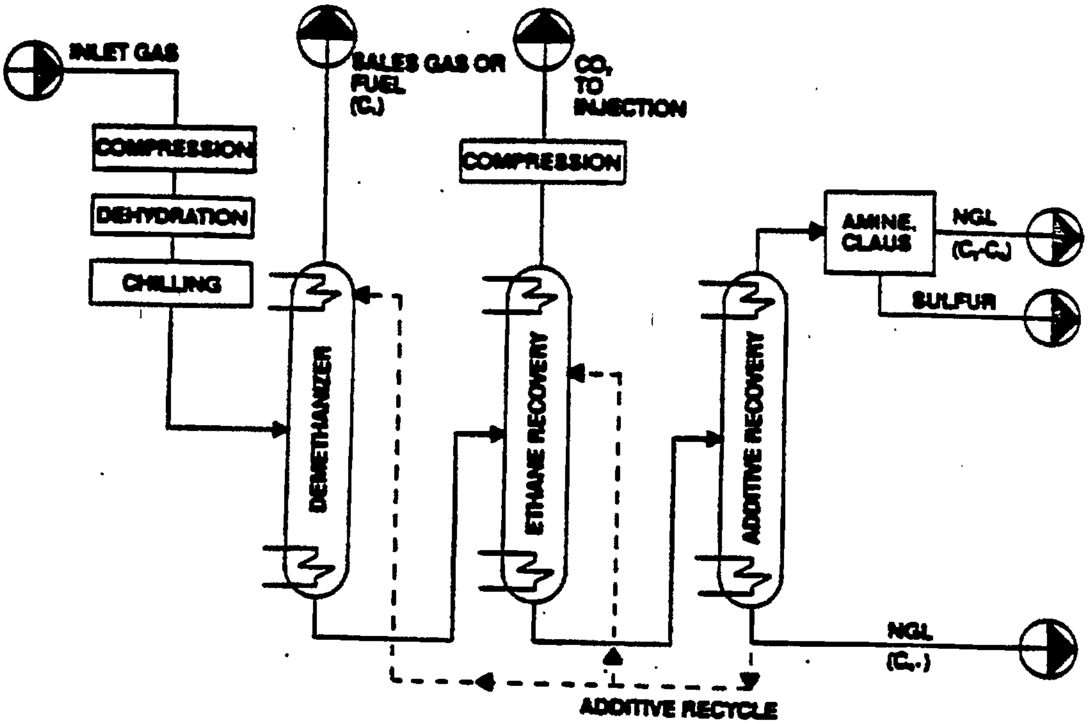

FIGURE 4-1 3 COLUMN RYAN/HOLMES PROCESS (SOURCE REF. #2)

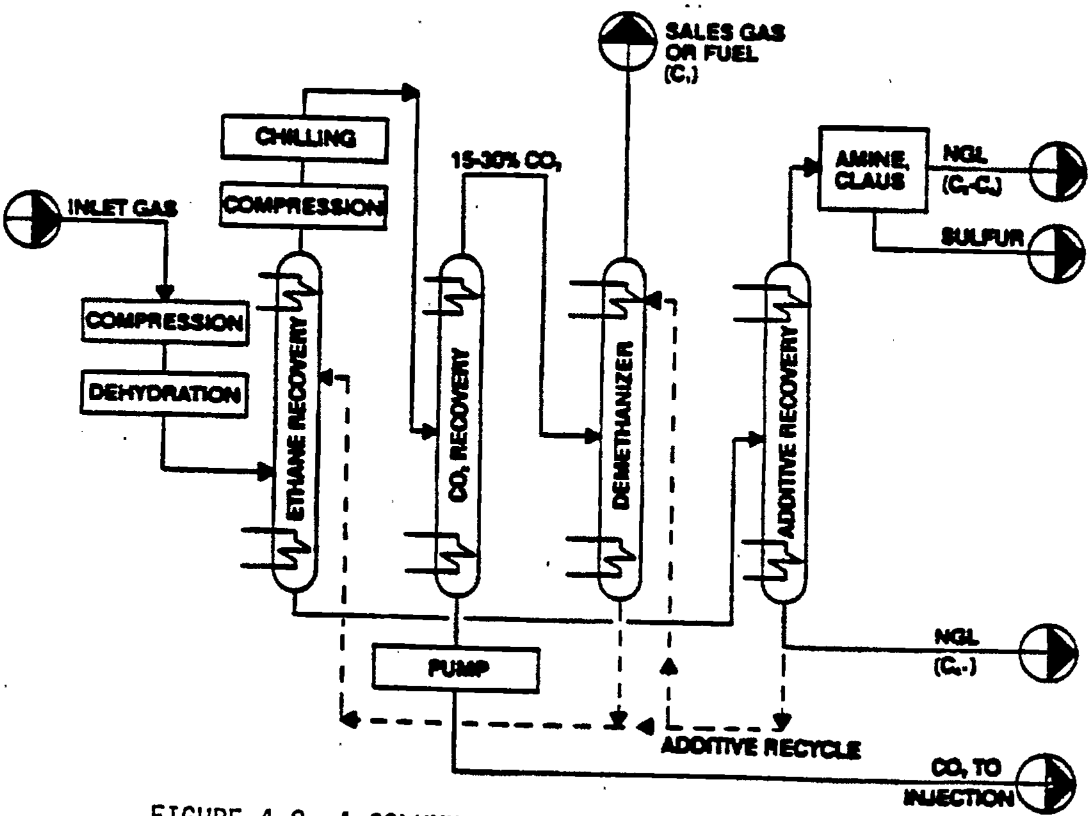

FIGURE 4-2 4 COLUMN RYAN/HOLMES PROCESS (SOURCE REF.#2)

K-F 1026-1 (1-78)

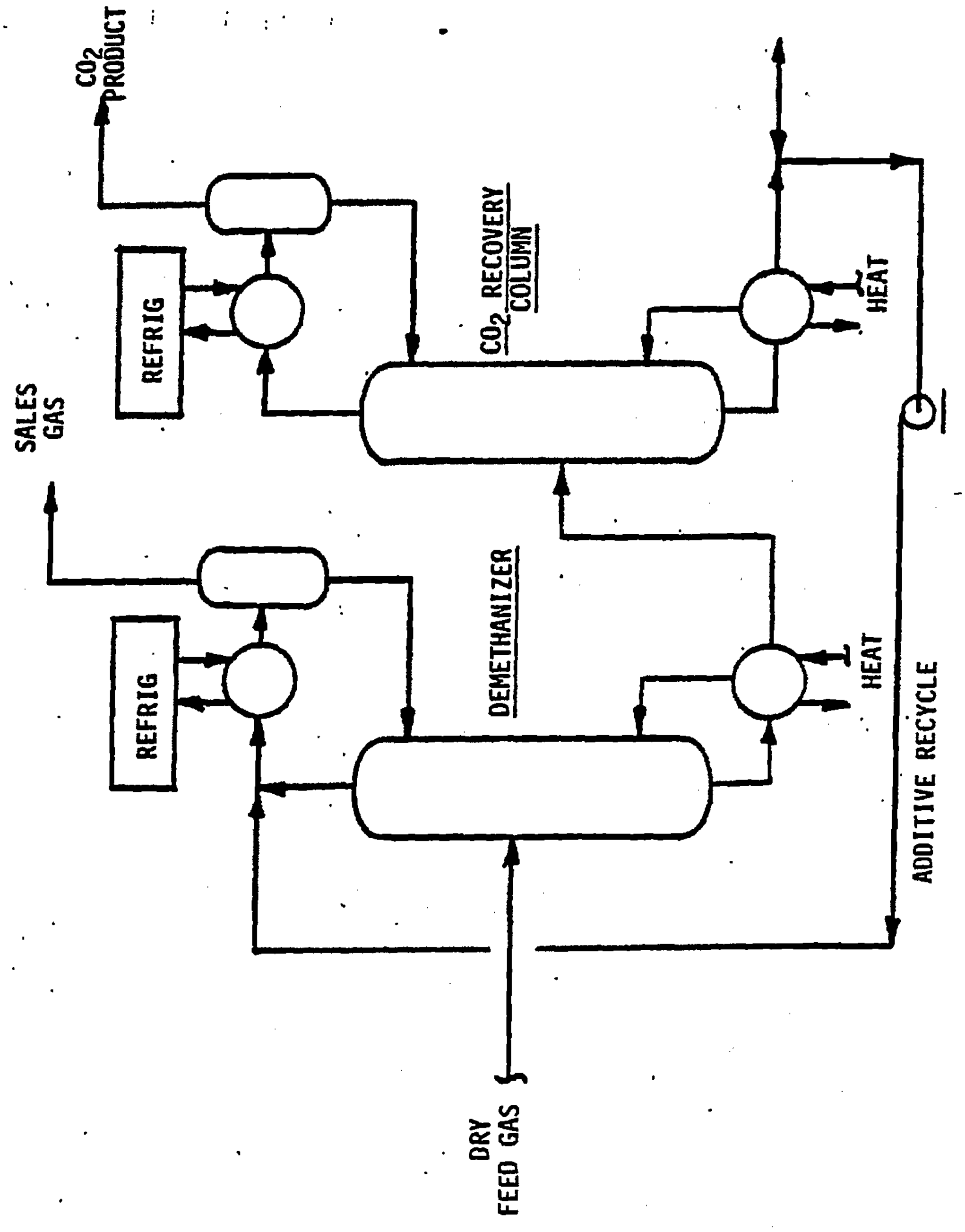

FIGURE 4-3
RYAN-HOLMES PROCESS FOR UPGRADING BTU CONTENT
(No H_2S Present)

NO.	REVISION DESCRIPTION	BY	CHK.	APPR.	DATE
ISSUED FOR FABRICATION		ISSUED FOR CONSTRUCTION			

SCALE:
DRWN:
CHK:
APPR:
DATE:

The M. W. Kellogg Company
GAS RESEARCH INSTITUTE

CLASS	AREA	JOB NO.	DRAWING NO.	REV
		6440-10		

K-F 1026-1 (1-78)

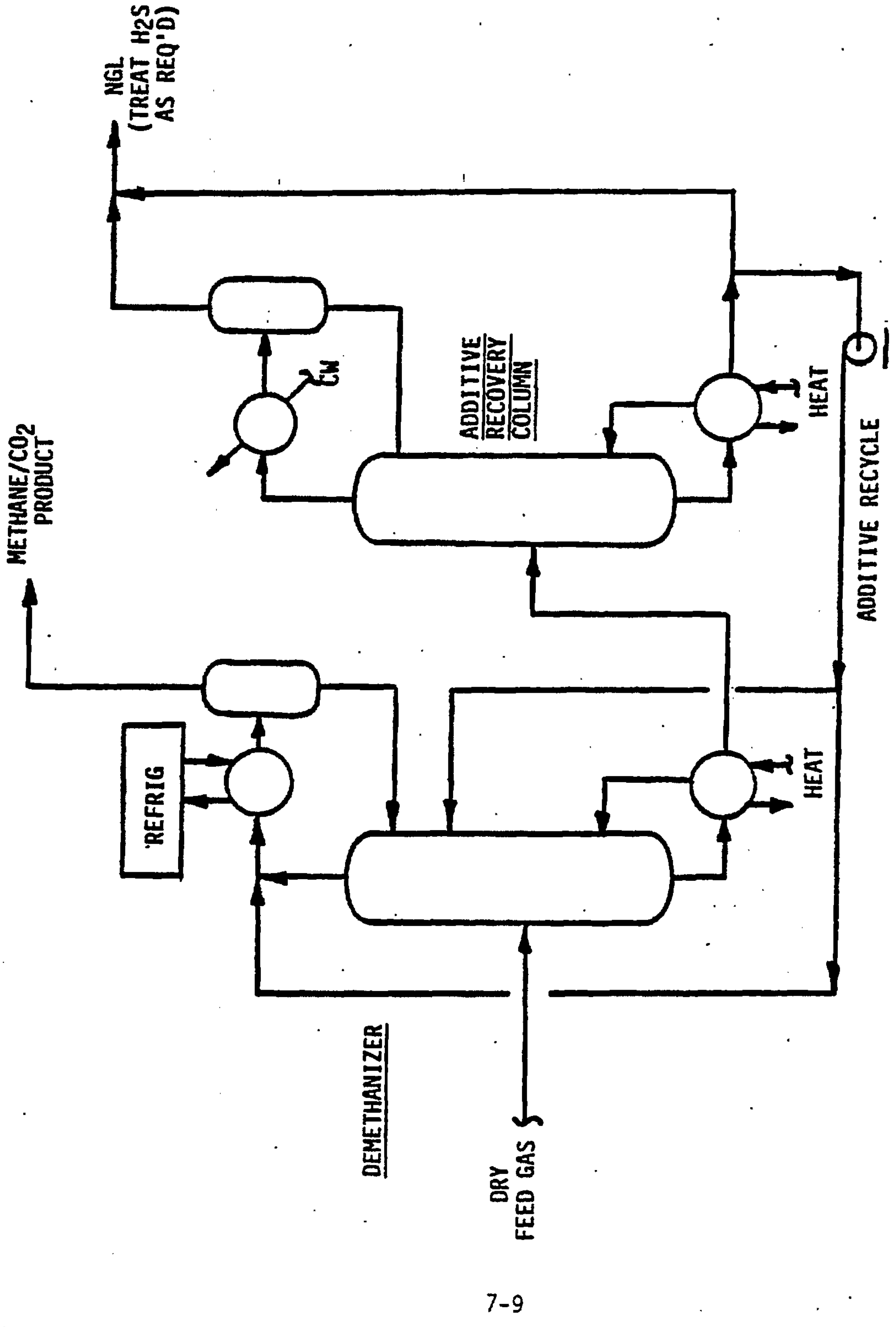

FIGURE 4-4
RYAN-HOLMES PROCESS
FOR NGL RECOVERY

NO. | REVISION DESCRIPTION | BY | CHK. | APPR. | DATE

ISSUED FOR FABRICATION

ISSUED FOR CONSTRUCTION

SCALE:
DRWN:
CHK:
APPR:
DATE:

The M. W. Kellogg Company

GAS RESEARCH INSTITUTE

CLASS	AREA	JOB NO.	DRAWING NO.	REV
		6440-10		

5.0 COMMERCIAL DESIGN DATA

- Enhanced Oil Recovery Gas Separation Plant (Refer to Figure 4-1)

Moles/Hr:

	Feed Gas	Fuel Gas Product	CO_2 to Injection	NGL Product	% Recovery
CO_2	793.3	4.2	715.1	20.0	
H_2	5.0	-	4 ppm	5.0	
CH_4	214.2	199.2	15.0	-	
C_2H_6	110.9	-	8.0	102.9	92.8
C_3H_8	94.8	0.1	.3	94.4	99.6
C_4	116.8	0.7	-	116.1	99.4
N_2	3.1	3.1	-	-	-
TOTAL	1284.1	207.3	738.4	338.4	

- High CO_2 Content Gas Separation

Basis:
Pipeline Sales Gas, 1000 psig, 2% CO_2
Product CO_2, 2000 psig
Feed 2000 psig

Feed:	CH_4, MMSCFD	27
	CO_2, MMSCFD	73
	TOTAL	100
Product:	Sales Gas, MMSCFD	22.5
	CO_2 Product, MMSCFD	73.4
	Fuel Product, MMSCFD	4.1
	TOTAL	100.0

(Source Ref. 1)

6.0 ADVANTAGES AND DISADVANTAGES

- Advantages

  - Can produce pipeline quality methane, with H_2S less than 4 ppm and CO_2 less 1%.
  - Can produce a Claus plant feed stream with a high H_2S/CO_2 ratio.
  - Can produce a CO_2 stream with less than 100 ppm H_2S.
  - Up to 98% recovery of ethane can be achieved.
  - Low loss of additive agent, C_4^+, to products, e.g., approximately 1%.
  - Flexibility to handle widely-varying flow rates and compositions of feedstock.
  - Use of expander/compressor sets not required.
  - High recovery of CH_4^+.

- Disadvantages

  - If the overhead condenser is operated at approximately -60°F instead of approximately -120°F, the additive agent/methane molar ratio increases from 0.2 to 4.0.

The Ryan-Holmes processes are proven only for natural gas and NGL applications. Very little, if any, investigation has been devoted to processing of synthesis gas. Representatives of Koch have indicated concern that significant concentrations of hydrogen would not be easily handled by the processes.

7.0 COMMERCIAL INSTALLATIONS (Reference 2)

- Seminole Unit operated by Amerada Hess
- Willard Unit, Wasson Field, operated by ARCO Oil & Gas.
- GMK Smith Field operated by Mobil Producing Texas and New Mexico Inc.
- Wasson Denver Unit operated by Shell Oil Company.

8.0 REFERENCES

1. Holmes, A. S., J. M. Ryan, B. C. Price, and Styring; "Pilot Tests Prove Ryan/Holmes Cryogenic Acid Gas/Hydrocarbon Separations"; Paper presented at 61st annual GPA Convention. March 1982, Dallas,Texas.

2. Ryan, J. M., and F. W. Schaffert; "The Ryan/Holmes Technology - An Economical Route for CO_2 and Liquids Recovery"; Paper presented at the 1984 AIchE Winter National Meeting, March 1984, Atlanta, Georgia.

8.0 COMPARISON OF PERFORMANCE/DESIGN PARAMETERS

The GRI/Advisors Planning and Strategy (GAPS) Committee was established to develop a plan for guiding research in the areas of fossil fuel gasification and downstream gas processing. As part of the work, the committee has developed a procedure for evaluating acid gas removal processes by setting up performance criteria to evaluate processes. This allows the identification of specific advantages and disadvantages of various processes and to establish research goals for process improvement and new process development. The "Musts" in acid gas removal technology are shown in Table 8-1. The technical criteria and standards developed for AGR technology appear in Table 8-2. A brief description and explanation of the same is provided where appropriate. Table 8-3 summarizes the performance of the four (Selexol, Rectisol, Benfield and CNG) processes. The Ryan-Holmes process is not included in the evaluation, for the reason it has not yet been applied or proven to process synthesis gas from coal gasification units. All the data in Table 8-3 are extracted from the respective Status Summary reports and from the public sources. It should be noted that these data are based on current publicly available resources; as more data are developed or made available to the public by the licensors, this table could be updated. Footnotes at the end of the table are provided for additional clarification.

TABLE 8-1

'MUSTS'IN ACID GAS REMOVAL TECHNOLOGY SELECTION

- The acid gas removal technology being considered must:

1. Be capable of processing the typical gas from coal gasification to effect removal of at least 95% of the contained H2S and at least 98% of the contained CO2.

2. Be compatible with at least one type of shift/methanation system, i.e., coventional, combined or direct, and with at least one standard commercial sulfur recovery process.

3. Generate only residues which are disposable using available technology, i.e., solid residues suitable for landfill without major environmental control, liquid residues convertable to disposable effluents and gaseous residues convertible to ventable effluents.

4. Requires no highly exotic materials of construction.

5. Involves no solvent or process reagent which is regarded as highly toxic or highly carcinogenic.

6. Be developed such that the basic concept is confirmed.

TABLE 8-2

TECHNICAL CRITERIA AND STANDARDS FOR ACID GAS REMOVAL TECHNOLOGY

CRITERION	DESCRIPTION	STANDARD	EXPLANATION
1.0 Feedgas Processing Capability (Concentration of components, vol%):			
1.1 CO2	Ability to process feed gas having a widely varying concentration of constituents.	5 - 60	The process must be capable of treating the raw gas from a variety of gasification technologies.
1.2 H2S		0.2 - 1.3	
1.3 COS		100-500 ppmv	
1.4 C2-C4 Gases		Up to 1.2	
1.5 C5+ hydrocarbons		Up to 0.2	
1.6 Water		0.3 - 0.5	
1.7 Other Contaminants		Up to 50 ppm NH3/HCN	
2.0 Separation Capability:			
2.1 H2S in product gas	Ability to remove CO2 and sulfur compounds (H2S and COS) for compatibility with sulfur recovery processes and in compliance with environmental regulations.	1 - 5 ppm	
2.2 CO2 in product gas		1 vol%	
2.3 H2S in acid gas- (selective AGR)		25% minimum	For selective AGR processes to meet Claus feed specification.
2.4 H2S in acid gas- (non-selective AGR)		1% minimum	For non-selective AGR processes.
2.5 Sulfur in vent gas		250ppm as SO2	
2.6 CH4 lost from product gas		< 1.0 %	As percentage of CH4 in feed gas.
2.7 C2-C4 hydrocarbons lost from product		No losses	Index vs Selexol
2.8 COS removal efficiency		maximum	Ratio of COS in H2S-rich gas to COS in AGR feed.
3.0 Process Efficiency (percent):			
3.1 Product Gas Loss	100 * (1- HHV product gas/ HHV sulfur-free feed gas)	< 1 %	Defines the product gas losses to H2S and CO2 rich streams.
3.2 Energy Consumption	Utility input(MM BTU)/HHV of feed gas.	< 5 %	Utility requirements—include steam and electric power, which must be converted to an equivalent heat rate.
4.0 By-product Production:			
4.1 CO2 Impurity		< 1 %	High purity CO2 could be sold as a by-product;also minimize product loss.
4.2 CO2 Recovery		> 75 %	

TABLE 8-2

TECHNICAL CRITERIA AND STANDARDS FOR ACID GAS REMOVAL TECHNOLOGY

(continued)

CRITERION	DESCRIPTION	STANDARD	EXPLANATION
5.0 Operating Conditions (Absorption/Regeneration):			
5.1 Absorption Temperature		100 F	Process with minimum feed gas cooling and no refrigeration requirement is preferable.
5.2 Absorption pressure		400 psig	Process should be capable of operating at the existing system pressure without the need of letdown or compression. Std. based on typical operating pressure of Lurgi,Westinghouse and BGC/Lurgi gasifiers.
5.3 Regeneration Temperature		100 F	Regeneration temp closest to the absorption temp is preferable.Process with no heat input or refrigeration requirement is preferred.
5.4 Regeneration Pressure		20 psig	Regeneration pressure should allow integration with sulfur recovery process without any additional compression Higher regeneration pressure is preferred.
6.0 Solvent Characterization:			
6.1 Vaporization losses	Negligible vaporization losses.	< 20% /yr.	As percentage of in-process inventory (Basis: Selexol)
6.2 Degradation losses	Negligible degradation losses.	< 1% /yr.	
6.3 Regenerability	Regenerable by more than two techniques.	> 2 Methods	Stripping methods: steam stripping, N2/Air stripping and pressure let down.
6.4 Safety in Solvent Handling	Solvent should be non-toxic, non-corrosive and inflammable.	Yes	Index relative to Selexol.

TABLE 8-2

TECHNICAL CRITERIA AND STANDARDS FOR ACID GAS REMOVAL TECHNOLOGY

(continued)

CRITERION	DESCRIPTION	STANDARD	EXPLANATION
7.0 Integrability with upstream and downstream units:			
7.1 Feed pretreatment	Number of steps(COS hydrolysis, carbon adsorption, mol. sieves)required to prepare feed suitable for acid gas removal unit.	No extra steps	
7.2 Compatibility with gas conversion processes	Process must be compatible with conventional Shift/Methanation, combined Shift/Methanation and direct methanation processes.	2 +	Compatible with at least two types of gas conversion systems.
7.3 Integrability with Claus sulfur recovery.	Number of extra steps required before feeding acid gas to the sulfur recovery unit.	No extra steps	
7.4 Integrability with Stretford sulfur recovery unit.	Number of extra steps required before feeding acid gas to the sulfur recovery unit.	No extra steps	
8.0 Process Techniques:			
8.1 Equipment Availability	Use of equipment which is readily available or requires minimal extrapolation from proven ranges.	No Extrapolation	
9.0 Materials of Construction:			
9.1 Availability	Materials of construction available at reasonable cost.	Standard Materials	
10.0 Complexity:			
10.1 Absorption steps	Number of towers in series must not exceed number of gas streams separated minus one.	2 Maximum	Two for selective process and one for non-selective process.
10.2 Regeneration Steps	Same as above.	2 Maximum	
10.3 Gas Recycles	Number of gas recycles should not exceed the number of removal steps.	2 Maximum	
10.4 Mechanical Complexity	No special internals,mechanical devices or moving parts required. Trays and packings considered as standard internals.	None	Index relative to Selexol.

TABLE 8-2

TECHNICAL CRITERIA AND STANDARDS FOR ACID GAS REMOVAL TECHNOLOGY

(continued)

CRITERION	DESCRIPTION	STANDARD	EXPLANATION
11.0 Controllability :			
11.1 Control System	Use existing control techniques or only minimal extrapolation.	Yes	Index relative to Selexol.
11.2 Turn down	Fraction of normal rate to which flows can be reduced while producing product and acid gases to specification.	To 50 %	
12.0 Reliability:			
12.1 Standby Requirements	Need for spare equipment (other than normal practise)	None	
12.2 Consequence of Failure		Minimum	Likelihood that damage would occur to other units due to a failure of AGR system.
12.3 Maintenance Extent	Need for maintenance	Minimum	
12.4 Overdesign Requirements	Necessity to overdesign equipment for coal composition variations.	None	
13.0 Environmental Considerations:			
13.1 Solid Effluents 13.2 Liquid Effluents 13.3 Gaseous Effluents	Number of steps required to dispose the effluents in environmentally acceptable manner.	None None None	

TABLE 8-3

PERFORMANCE OF ACID GAS REMOVAL TECHNOLOGIES VS. CRITERIA

CRITERIA AND SUBCRITERIA	EXPLANATION	SELEXOL	RECTISOL	BENFIELD	CNG
1.0 FEEDGAS PROCESSING CAPABILITY					
1.1 CO2	Process wide	No limitation	No limitation	No limitation	16 vol% Minimum @ 450 psig
1.2 H2S	range of concen-	0.2 to 1.3 vol %	0.2 to 1.3 vol %	0.2 to 1.3 vol %	0.2 to 1.3 vol %
1.3 COS	trations of each	10 - 20 ppmv	Up to 500 ppmv	500 ppmv	500 ppmv
1.4 C2-C4 gases	species.	Up to 1.2 vol %	Up to 1.2 vol %	> 1.2 VOL %	> 1.2 VOL %
1.5 C5+ hydrocarbons		Prewash req'd.	Prewash req'd.	Independent Prewash	None Tolerated
1.6 Water		0.3 to 0.5 vol %	0.3 to 0.5 vol %	0.5 vol %	None Tolerated
1.7 Other Contaminants		Up to 50 ppmv HCN/NH3	Up to 50 ppmv HCN/NH3	HCN produces non-regenerable salts	?????
2.0 SEPARATION CAPABILITY					
2.1 H2S in Product Gas	Remove CO2 and	1 to 5 ppm	Less than 1 ppm	1 ppm (Hipure)	1 ppm
2.2 CO2 in Product Gas	sulfur compounds	1 vol %	10 ppm	< 1 vol %	0.1 to 1.5 vol %
2.3 H2S in Acid Gas (S)	for compatability	25 to 30 %	25 to 30 %	2-5 vol %	50-75 %
2.4 H2S in Acid Gas (NS)	with sulfur re-	3-5 vol %	3-5 vol %	1-2 vol %	50-75 %
2.5 Sulfur in Vent Gas	covery processes	< 250 ppmv	< 250 ppmv	No Vent	< 1 ppm
2.6 CH4 Loss	As % of CH4 in feed gas	0.4 - 0.5 %	0.8 - 1.0 %	< 1 %	2 %
2.7 C2-C4 Losses	Index	1.0	0.9	1.1	0.8
2.8 COS Removal Effic	In H2S rich gas vs.AGR feed	10 to 20 %	Over 90 %	70-90 %	99 + %

S Selective
NS Non-selective

TABLE 8-3(Cont'd)

PERFORMANCE OF ACID GAS REMOVAL TECHNOLOGIES VS.CRITERIA

CRITERIA AND SUBCRITERIA	EXPLANATION	SELEXOL	RECTISOL	BENFIELD	CNG
3.0 PROCESS EFFICIENCY					
3.1 Product Gas Loss	(1-HHV product gas/HHV sulfur free feed gas)	4% (Lurgi) 2% (Westghs)	6.5% (Lurgi) 1.4% (Exxon)	1.7% (Westghs)	3.5% (Bigas)
3.2 Energy Consumption	Utility input/ HHV of feed gas	4.7 % (Westghs)	4.5% (Exxon) 3.4% (Lurgi)	13.9%(Westghs)	1.4% (Bigas)
4.0 BYPRODUCT PRODUCTION					
4.1 CO2 Impurity		1 % or less.	1 % or less.	Not Produced.	< 0.5 %
4.2 CO2 Recovery		75 - 85 %	50 - 70 %	Not Produced.	90 + %
5.0 OPERATING CONDITIONS					
5.1 Absorber Temperature	deg F	100 degF (H2S) 50 degF (CO2)	-50 to -70 degF (Solvent temp)	200-250 degF	-67 to -112 degF
5.2 Absorber Pressure	psig	400 psig	400 psig	400 psig	400 psig
5.3 Regenerator Temp	deg F	280 degF(H2S) 10-20 degF(CO2)	250 degF(H2S) 5 degF(CO2)	200-250 degF	-67 degF
5.4 Regenerator Press	psig	20 psig(H2S) 2 - 5 psig(CO2)	5 - 10 psig	2-5 psig	> 75 psig
6.0 SOLVENT CHARACTER					
6.1 Vaporization Loss	% of in-process inventory	20 % per year	240-700 % per yr	30 % per year	?
6.2 Degradation Loss		Negligible.	Negligible.	> 1% per year	Negligible
6.3 Regenerability	Methods	Flash,Steam,N2	Flash,Steam,N2	Steam Stripping	Flash,N2
6.4 Solvent Safety	Index	1.0	0.8	1.1	0.9

TABLE 8-3(Cont'd)

PERFORMANCE OF ACID GAS REMOVAL TECHNOLOGIES VS. CRITERIA

CRITERIA AND SUBCRITERIA	EXPLANATION	SELEXOL	RECTISOL	BENFIELD	CNG
7.0 INTEGRATION WITH OTHER PROCESS AREAS					
7.1 Feed Pretreatment	Extra steps	COS Hydrolysis & Oil Removal	Dehydration & Oil Removal	Oil Removal & COS Hydrolysis	Dehydration and Oil Removal
7.2 Compatability with Gas Conversion	Different type of processes	3 Types	3 Types	3 Types	3 Types
7.3 Integrability with Claus Unit	Steps required before feeding	None.	None.	One	None.
7.4 Integrability with Stretford Unit	Steps required before feeding	None.	None.	None.	None.
8.0 PROCESS TECHNIQUES					
8.1 Equip't Available	Standard vessels	No extrapolation	No extrapolation	No extrapolation	Will require extrapolation
9.0 MATERIALS OF CONSTRUCTION					
9.1 Availability	None exotic	Standard mat'ls	Standard mat'ls	Standard mat'ls	Standard Mat'ls
10.0 COMPLEXITY					
10.1 Absorption Steps		Two(S), One(NS)	Three(S), Two(NS)	Two(Hipure)	Two
10.2 Regeneration Steps		Two(S), One(NS)	Three	Two	Two
10.3 Gas Recycles		Two	One	None	Two
10.4 Mechanical Complexity	Index	1.0	0.9	1.1	0.9
11.0 CONTROLLABILITY					
11.1 Control System	Index	1.0	1.0	1.0	0.8
11.2 Turndown	% of full rate	50 %	50%	50 %	25-50 %

TABLE 8-3(Cont'd)

PERFORMANCE OF ACID GAS REMOVAL TECHNOLOGIES VS.CRITERIA

CRITERIA AND SUBCRITERIA	EXPLANATION	SELEXOL	RECTISOL	BENFIELD	CNG
12.0 RELIABILITY					
12.1 Standby Reqmnts	Active/total	None	None	None	Some
12.2 Consequence Of Failure	Risk involved	Minimum	Minimum	Minimum	Minimum
12.3 Maintenance Extent	Index	1.0	1.0	1.1	0.9
12.4 Overdesign Requirements		None	None	None.	None
13.0 ENVIRONMENTAL CONSIDERATIONS					
13.1 Solid Effluents	Extra steps	None	None	None.	None
13.2 Liquid Effluents	Extra steps	None	Purge stream	Spent solvent	None
13.3 Gaseous Effluents	Extra steps	None	None	None	None

CRITERIA	EXPLANATION
1.1	CO_2 IN FEED - Selexol, Rectisol and Benfield can process any CO_2 concentration. CNG requires a minimum CO_2 partial pressure of 75 psia (16% CO_2 at 450 psia) in the feed gas.
1.3	COS IN FEED - CNG and Rectisol remove COS with H_2S stream while Selexol removes it with the CO_2 stream. Benfield removes it by hydrolysis to H_2S.
1.5	C_5+ HYDROCARBONS - CNG cannot tolerate any C_5+'s due to freezing possibilities.
1.6	WATER IN FEED - Benfield can tolerate maximum water because of higher feed temperature.
2.1	H_2S IN PRODUCT GAS - Selexol can produce 1 ppm H_2S product gas with additional trays.
2.3	H_2S IN ACID GAS (SELECTIVE) - Selexol is expected to produce a somewhat less concentrated H_2S stream as compared to Rectisol for the same coal type.
2.4	H_2S IN ACID GAS (NON-SELECTIVE) - Non-selective Rectisol and Selexol produce an acid gas which is relatively richer in H_2S as compared to Benfield.
2.5	SULFUR IN VENT GAS - Vent gas from Selexol is likely to have a higher SO_2 content as compared to Rectisol since COS is removed with the CO_2 stream in Selexol.
2.8	COS REMOVAL EFFICIENCY - Selexol is least effective in removing COS with H_2S stream.
3.1	PRODUCT GAS LOSS - This is a function of the raw gas composition (i.e., the type of gasifier). In general, the losses will increase along with the methane content of the raw gas.
3.2	ENERGY CONSUMPTION - This is also a function of the feed composition because the volume of gas to be treated from a fixed bed gasifier will be less than that from an entrained bed gasifier.
4.1	CO_2 BYPRODUCT PURITY - Selexol CO_2 stream will contain COS.
5.2	ABSORBER PRESSURE - CNG requires a certain minimum CO_2 partial pressure in the feed, which may necessitate feed compression.

6.1 VAPORIZATION LOSS - For CNG, sufficient information is not available but it is known that the solvent for CO_2 removal is a heavier hydrocarbon.

7.2 COMPATIBILITY WITH GAS CONVERSION PROCESSES - All processes can be used in conjunction with any of the gas conversion processes (conventional methanation, combined shift/methanation and direct methanation) but the sulfur content to zinc oxide beds is lowest for CNG and higher but about the same for others.

7.3 INTEGRABILITY WITH CLAUS UNIT - CNG produces highest concentration H_2S stream and because Benfield is a non-selective process it requires further treatment to obtain a sufficiently concentrated H_2S-rich stream.

8.1 EQUIPMENT AVAILABILITY - The scale-up flexibility of the crystallizer for the CNG process is undefined.

10.2 REGENERATION STEPS - Rectisol has three steps due to the presence of methanol column regenerator.

10.4 MECHANICAL COMPLEXITY - Benfield is considered less complex because of no recycles, solvent coolers, etc. Rectisol is more complex than Selexol due to refrigeration requirements. CNG is the most complex because it pumps a solid slurry and due to the crystallizer.

11.1 CONTROL SYSTEM - For CNG, control of crystallizer may require special control techniques.

12.1 STANDBY REQUIREMENTS - CNG was judged to require some parallel equipment such as the crystallizer.

12.4 OVERDESIGN REQUIREMENTS - This defines the need for overdesign of equipment to meet such variations as sulfur in coal. CNG is judged to respond most favorably to these variations.

13.2 LIQUID EFFLUENTS - Rectisol generates a water/methanol purge stream which must be treated. Benfield generates a spent solvent stream which must be disposed.