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Equipment Design and Cost Estimation for Small Modular Biomass Systems, Synthesis Gas Cleanup, and Oxygen Separation Equipment

Task 1: Cost Estimates of Small Modular Systems

Nexant Inc.
San Francisco, California

Subcontract Report
NREL/SR-510-39943
May 2006

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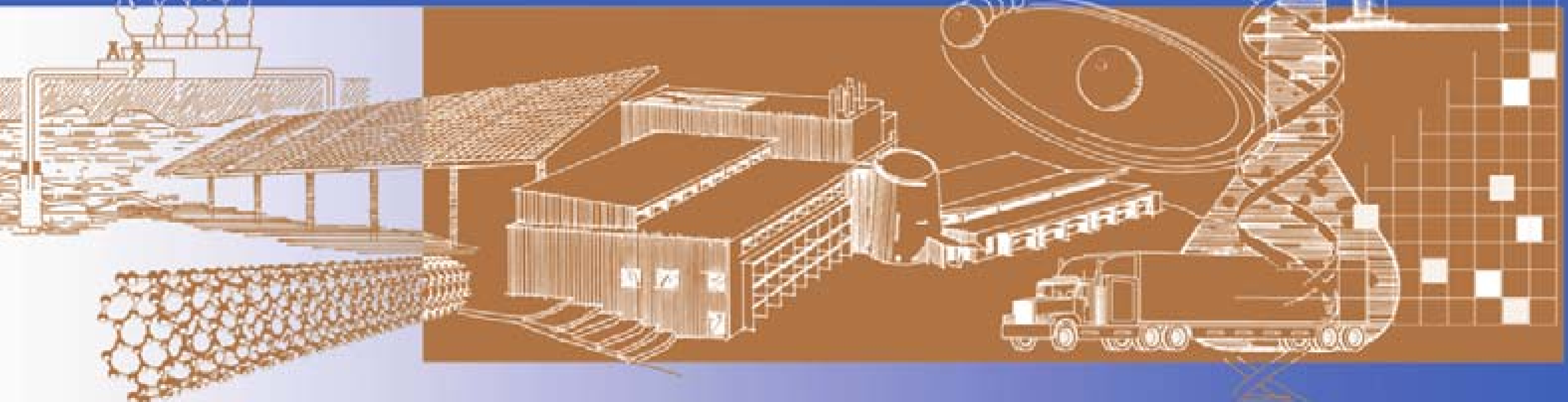

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NREL Technical Monitor: Kelly Ibsen
Prepared under Subcontract No. ACO-5-44027

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Executive Summary

This deliverable is the Final Report for Task 1, Cost Estimates of Small Modular Systems, as part of NREL Award ACO-5-44027, "Equipment Design and Cost Estimation for Small Modular Biomass Systems, Synthesis Gas Cleanup and Oxygen Separation Equipment". Subtask 1.1 looked into processes and technologies that have been commercially built at both large and small scales, with three technologies, Fluidized Catalytic Cracking (FCC) of refinery gas oil, Steam Methane Reforming (SMR) of Natural Gas, and Natural Gas Liquids (NGL) Expanders, chosen for further investigation. These technologies were chosen due to their applicability relative to other technologies being considered by NREL for future commercial applications, such as indirect gasification and fluidized bed tar cracking. Research in this subject is driven by an interest in the impact that scaling has on the cost and major process unit designs for commercial technologies. Conclusions from the evaluations performed could be applied to other technologies being considered for modular or skid-mounted applications.

Subtask 1.2.1 presented PFDs and preliminary material/energy balances for each of the designs. Subtasks 1.2.2 through 1.2.4 built upon the information presented in Subtask 1.2.1 by providing equipment specification sheets, capital and installation costs, cost breakdowns for individual pieces of equipment, and operating costs. The following deliverable provides detail for large and small scale designs of each technology.

For each technology, a basis was established for throughput and cost at each scale. Results from the technical analysis can be seen in Tables A through C:

TABLE A TOTAL FCC PROJECT COSTS

	Large Scale Plant	Small Scale Plant
Gas Oil Feed Rate (BPSD)	62,000	2,500
Reactor/Regen TPI ($MM)	102.6	11.0
Total Project Investment ($MM)	191.6	20.5
Total Investment/BBL Feed ($)	~3,000	~8,200

TABLE B TOTAL STEAM METHANE REFORMER PROJECT COSTS

	Large Scale Plant	Small Scale Plant
Natural Gas Feed Rate (MMSCFD)	167.8	0.024
Installed Cost ($MM)	244.0	0.26
Total Plant Investment ($MM)	368.5	0.39
Total Investment/SCFD Feed ($)	2.2	16.2

TABLE C TOTAL NATURAL GAS EXPANDER PROJECT COSTS

	Large Scale Plant	Small Scale Plant
Natural Gas Feed Rate (MMSCFD)	400	75
Installed Cost ($MM)	46.4	12.0
Total Plant Investment ($MM)	62.4	16.1
Total Investment/SCFD Feed ($)	0.16	0.21

Although the technologies considered and the relative ranges between the large and small scale designs were all different, a common relationship existed between the large and small scale designs that could be applied to other technologies. Using a simple power relationship defined by the following,

$$\text{Small Scale Cost} = \text{Large Scale Cost} * \left(\frac{\text{Small Scale Capacity}}{\text{Large Scale Capacity}} \right)^{0.7}$$

it was determined that the relationship between each case could be predicted with some degree of accuracy. Table D below shows the size and cost ratios for each technology, along with the predictive capacity of using a power relationship.

TABLE D DESIGN RATIOS AND PREDICTED COSTS, ALL TECHNOLOGIES

	FCC	SMR	NG Expander
Large/Small Scale Size Ratio (Feedstock)	25	7,000	5.3
Large/Small Scale Cost Ratio	9.4	1,000	3.8
Size Ratio/Cost Ratio	2.7	7	1.4
Predicted Small Cost, Power Basis ($MM)	20.2	0.75	19.3
Actual Small Cost, Design Basis ($MM)	20.5	0.39	16.1
Percent Difference	1.5%	92.3%	19.8%

Using a power basis provided a cost estimate well-within an order of magnitude acceptability for each technology, and, for some, within a level of acceptability (+/- 30%) for this level of detail. Although applying a power relationship to SMR reforming technology had the worst predictive capacity, the resulting estimate is still very close on an absolute basis. It was expected that this technology would have the most prediction difficulty because of the considerable difference in size relative to the other technologies chosen.

A few general conclusions can be made regarding the cost relationship as technologies are scaled:

- The power relationship that should be used depends largely on the size range that is being considered. For example, a very large difference in sizes (SMR technology) will usually be much more difficult to accurately predict than smaller differences.
- Economies of scale are usually greater when technologies contain a novel or expensive main piece of equipment, such as the catalytic furnace in an SMR unit or a reactor/regenerator unit

in a FCC design. Established technologies, such as refrigeration and fractionation in the NG Expander designs, are more easily predicted as to their impact when sizes change.

- Many technologies have a breakpoint where the type of design changes due to size factors. This is true for all technologies considered in this study, as well as for many other technologies such as oxygen production units. As designs get close to this technology breakpoint, the capital cost per unit of feed changes considerably.

Introduction and Methodology

This study evaluated chemical process designs that have been successfully scaled for large and small scale commercial applications. The goal was to provide additional insight into how the cost and design of mature commercial technologies scale as the overall unit throughput changes. To accomplish this, a list of process technologies was paired down to three for further evaluation: Fluidized Catalytic Cracking of refinery gas oil, Steam Methane Reforming of natural gas to hydrogen, and Natural Gas Liquids Expansion to separate the major components of natural gas. A list and description of all technologies considered for evaluation can be seen in Appendix A.

Complete designs for each technology on a large and small basis included process flow diagrams, equipment lists, stream properties, operating costs, and installation factors. After the designs were completed, comparisons were made between the different scales to determine the cost impacts. Finally, conclusions were drawn for how general scaling approximations can be used regardless of the technology.

The methodology for designing large and small schemes for each process was somewhat different. For the natural gas reforming and expansion technologies, the processes involved are non-proprietary. The equipment involved is standard to any chemical process industry: pumps, heat exchangers, distillation columns, pressure swing absorbers, and compressors. Based on the plant sizes and the range of plant information desired by NREL, complete theoretical facilities were designed by the project team. Information on all process flows, equipment sizes, energy requirements, and plant costs are presented in this deliverable.

FCC units, however, are a proprietary technology with specific detailed information unavailable to the project team. A number of licensers (UOP, ExxonMobil, KBR, and Stone and Webster) were contacted to obtain additional information about designs of FCC units at both large and small scales. This information was supplemented from literature sources and previous FCC studies worked on by a major EPC vendor. Per NREL guidelines, the area of the FCC unit focused on by the project team is the reactor/regenerator section. Process design information for this area of the plant was obtained from both licensers and literature sources for units already constructed at large and small scales. In addition, scaling information and cost curves representing a broad range of sizes have been obtained to check the results. Typical stream conditions for both units have been estimated from vendor information and literature data. Finally, ICARUS model runs, using vendor and literature sizes, were run to confirm information supplied by other sources for plant costs.

1.1 INTRODUCTION AND METHODOLOGY

Data was collected from publications, licensors, previous in-house designs, and owner/operators of Fluid Catalytic Cracking (FCC) units. The focus of this research was on the capital and operating costs of large and small scale FCC reactor/regenerators. The product recovery, gas separation, and flue gas treating sections are not included. Since this study includes both large and small scale designs, two process flow diagrams (PFDs) have been developed to reflect the type of configuration which is considered most economic for the unit size. This has led to two different designs:

- A "side-by-side" configuration of the reactor and regenerator, generally applicable to large scale units processing 20,000 or more barrels per stream day of fresh feed. This was used for the large scale unit, with a feed rate of 62,000 barrels per stream day (BPSD).
- A "stacked" configuration of the reactor and regenerator, generally applicable to small scale units processing less than 20,000 BPSD. This was used for the small scale unit, with a feed rate of 2,500 BPSD.

The equipment shown in these diagrams are based on the published designs of UOP. The costs established using the UOP design is representative of typical FCCs. All the major licensors confirmed that units of the same capacity designed by various licensors have similar capital costs. The variations are within the accuracy of this study of +/- 30%.

Design and cost information was obtained from three main sources:

- Direct Industry Contact:
  - Information for the large scale FCC design is based off a detailed design study performed by a major EPC vendor in the early 1990's.
  - Design information for portions of the small scale FCC comes directly from UOP data on an actual facility.
- Literature Sources:
  - A number of sources, footnoted in this report, were used for estimating operating costs and for providing checks on the industry derived numbers.
- Commercially Available Software:
  - ICARUS was used as a check against industry supplied numbers, and to provide estimates for pieces of equipment not costed by other sources.

1.2 TECHNOLOGY BACKGROUND

FCC technology was developed in the late 1930s and early 1940s due to the demand for aviation fuel in World War II. The first commercial FCC unit was brought on-stream in May 1942 at what is now known as the ExxonMobil Refinery in Baton Rouge, Louisiana. The FCC process converts gas oil and heavier streams in a refinery to lighter, more valuable products via high temperature catalytic cracking. Typical FCC products are propylene, butylenes, gasoline, coke, and dry gas (light ends, typically saturated hydrocarbons in the pentane and lighter range).

This technology is currently licensed by several companies. The main licensors include UOP LLC, EOM (ExxonMobil) / KBR (Kellogg Brown and Root) Alliance, and Stone and Webster Inc., each offering designs and features that enhance performance, reliability and flexibility of their units to accommodate a variety of feedstock and desired products.

There are more than 400 FCC units operating worldwide, from 2,000 to 135,000 BPSD of fresh feed. All the FCC units recently built are outside of the United States, with the largest being the Reliance FCC in Jamnagar, Gujarat, India. This unit was designed by UOP with an original nameplate design capacity of 135,000 BPSD. UOP reported that this unit achieved a record-breaking throughput of 180,000 BPSD in February 2003[1]. In the United States, the capacity of FCC units range from 2,400 to 110,000 BPSD of fresh feed. The smallest FCC unit is at Montana Refining in Great Falls, Montana. This unit was designed for 2,000 BPSD by UOP and built in the 1950s. The small-scale design in this report reflects process conditions for a unit of this size. Very few facilities in the US exist under 10,000 BPSD due to weak process economics; many facilities that existed in this range have either shutdown or expanded.

Over the years, with improvements in technology and increasing gasoline demand, most of the refiners have modified their FCC units to increase capacity. Several FCC units have been revamped to process more than 100,000 BPSD. According to the major licensors, the typical capacity of new grassroots FCC units in the United States is 40,000 to 60,000 BPSD. The large-scale design reflects the high end of this capacity, 60,000 BPSD.

Operating conditions of FCCs vary depending upon the feed, desired products, and mechanical limitations. The typical ranges of operational parameters are provided in a number of literature sources[2]. In addition, at the FCC Forum in May 2000, Engelhard (a major FCC catalyst manufacturer) presented the results of a FCC Benchmark study[3]. The study concluded that following are typical operating conditions of an “average” FCC in the United States:

- Processes 49,697 BPD of gas oil feed at an API gravity of 24.3 with 1,037 ppm nitrogen, 0.79 wt % sulfur, 0.9 wt % carbon residue, and 174°F aniline point.
- Operates at a reactor temperature of 978°F and a regenerator temperature of 1,299°F.
- Circulates 35.1 tons per minute of catalyst or 1.01 tons catalyst per barrel of feed.
- Uses catalyst at a rate of 6.2 tons per day or 0.25 pounds per barrel of feed.
- Produces 4.96 wt % coke and 3.23 wt % dry gas at a conversion level of 76 vol %.

---

[1] “Reliance FCC Unit Sets Capacity Record”, 4 March 2003, UOP Press Release, available at http://www.uop.com/objects/PR_RelCapacity.pdf

[2] The sources found best for this study were “Petroleum Refining—Technology and Economics” by J. H. Gary and G. E. Handwerk and “Handbook of Petroleum Refining Processes” by R.A. Meyers.

[3] Information presented by The FCC Network, Volume 4, June-July 2000, available at http://thefccnetwork.com/pdf/newsletters/newsletter4.pdf

- Yields 8.32 vol % propylene, 8.50 vol % butylenes, and 58.26 vol % gasoline with octane of 92.3 RON and 80.8 MON.

1.3 PROCESS DESCRIPTIONS AND DESIGN ASSUMPTIONS

The FCC process uses a catalyst in the form of very fine particles, averaging about 70 microns in size, which behave as a fluid when aerated with a vapor. The fluidized catalyst is circulated continuously between the reactor and the regenerator, and acts as a vehicle to transfer heat from the combustion of coke in the regenerator for heating the feed and for the endothermic cracking reaction. Until about 1965, most units were designed with a discrete dense-phase fluidized catalyst bed in the reactor vessel. The units were operated so most of the cracking occurred in the reactor bed. In the 1960’s, the development of more reactive zeolite catalysts was a major breakthrough in catalyst technology and led to process design changes to obtain high selectivity and control of the cracking reactions in the “riser” or transfer line. The reactor vessel serves as the disengaging space for the product vapor and the spent catalyst, as well as the housing for the cyclones.

The cracking reaction occurs in a vertical reactor riser in which a preheated liquid oil stream contacts hot catalyst from the regenerator. The oil vaporizes and cracks to lighter products as it moves up the riser and carries the catalyst powder along with it. Coke, a carbonaceous material having a low ratio of hydrogen to carbon (H/C) is a byproduct which deposits on the catalyst and reduces the catalyst activity. The spent catalyst and the converted products are separated in the reactor vessel. The cracked product vapors are sent to a fractionator and gas separation unit. Some of the hydrocarbons adsorbed on the spent catalyst are removed by steam stripping at the bottom of the reactor vessel. The spent catalyst passes to a separate vessel, the regenerator, where the coke is combusted to regenerate the catalyst. The hot, regenerated catalyst then passes to the bottom of the reactor riser, providing the heat needed in the reaction section.

Two basic types of FCC units in use today are the “side-by-side” type and the Orthflow or “stacked” type. The reactor and regenerator are separate vessels adjacent to each other for the “side-by-side” type, which is typical of designs by UOP, ExxonMobil, and Stone and Webster. The reactor is mounted on top of the regenerator for the “stacked” type, which is typical of designs by KBR.

Flow diagrams for both large and small scale cases can be seen in Figures 1-1 and 1-2, with corresponding stream properties in Tables 1-1 and 1-2. The design and cost estimates for each case are presented in the equipment list, which can be seen in Tables 1-3 and 1-4. The equipment list groups process equipment by the following categories: reactors, cyclones, columns/vessels/ tanks, and compressors.

TABLE 1-1 LARGE SCALE FCC STREAM PROPERTIES

PFD Stream No.	Description		Large Scale FCC
1	Feed	BPSD	62,000
2	Product Vapor	MMSCFD	100
3	Combustion Air	MMSCFD	146
4	Catalyst Circulation	Tons/min	50
5	Regenerator Flue Gas	MMSCFD	146

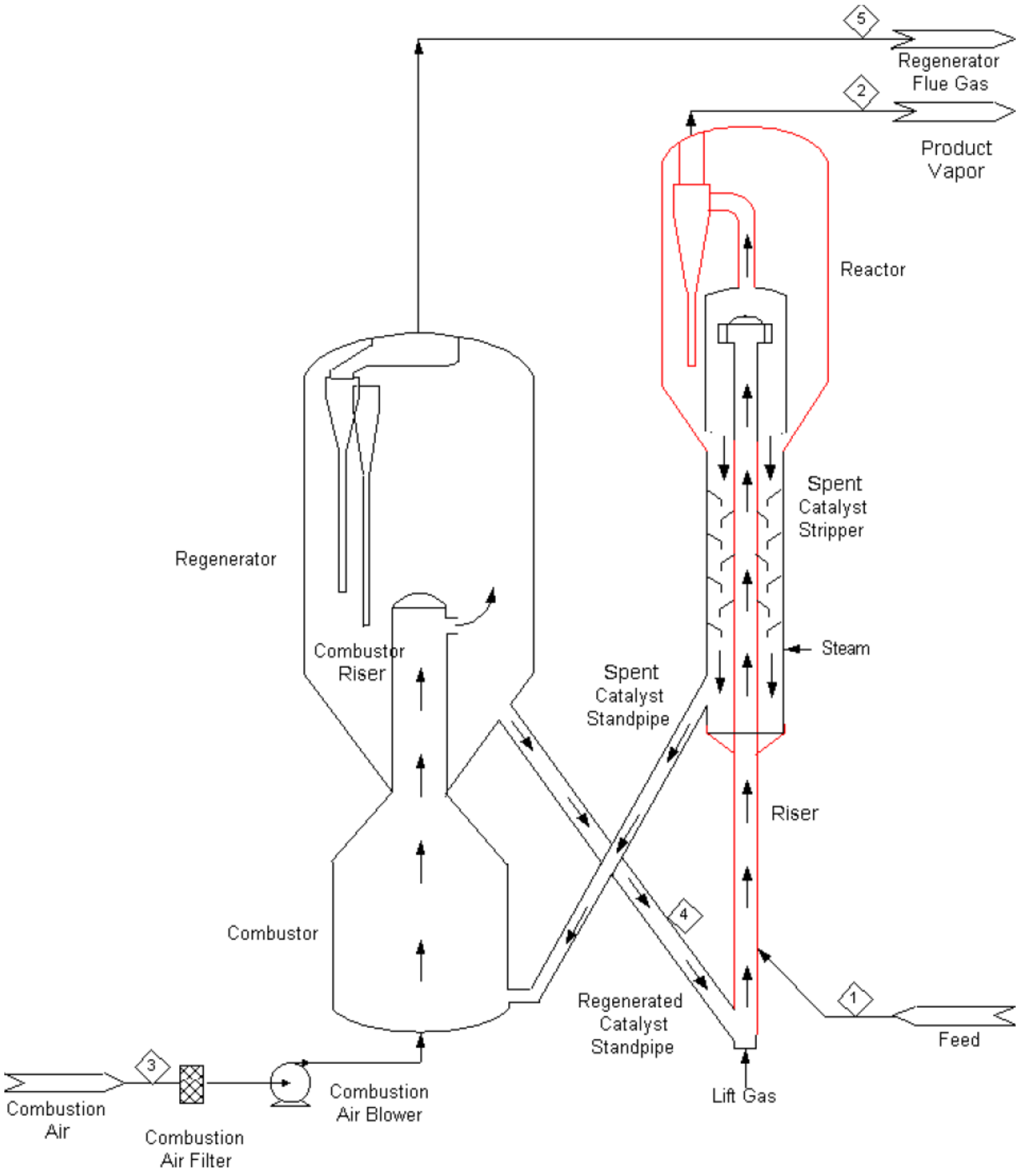

FIGURE 1-1 LARGE SCALE FCC DESIGN

TABLE 1-2 SMALL SCALE FCC STREAM PROPERTIES

PFD Stream No.	Description		Small Scale FCC
1	Feed	BPSD	2,500
2	Product Vapor	MMSCFD	4
3	Combustion Air	MMSCFD	6
4	Catalyst Circulation	Tons/min	2
5	Regenerator Flue Gas	MMSCFD	6

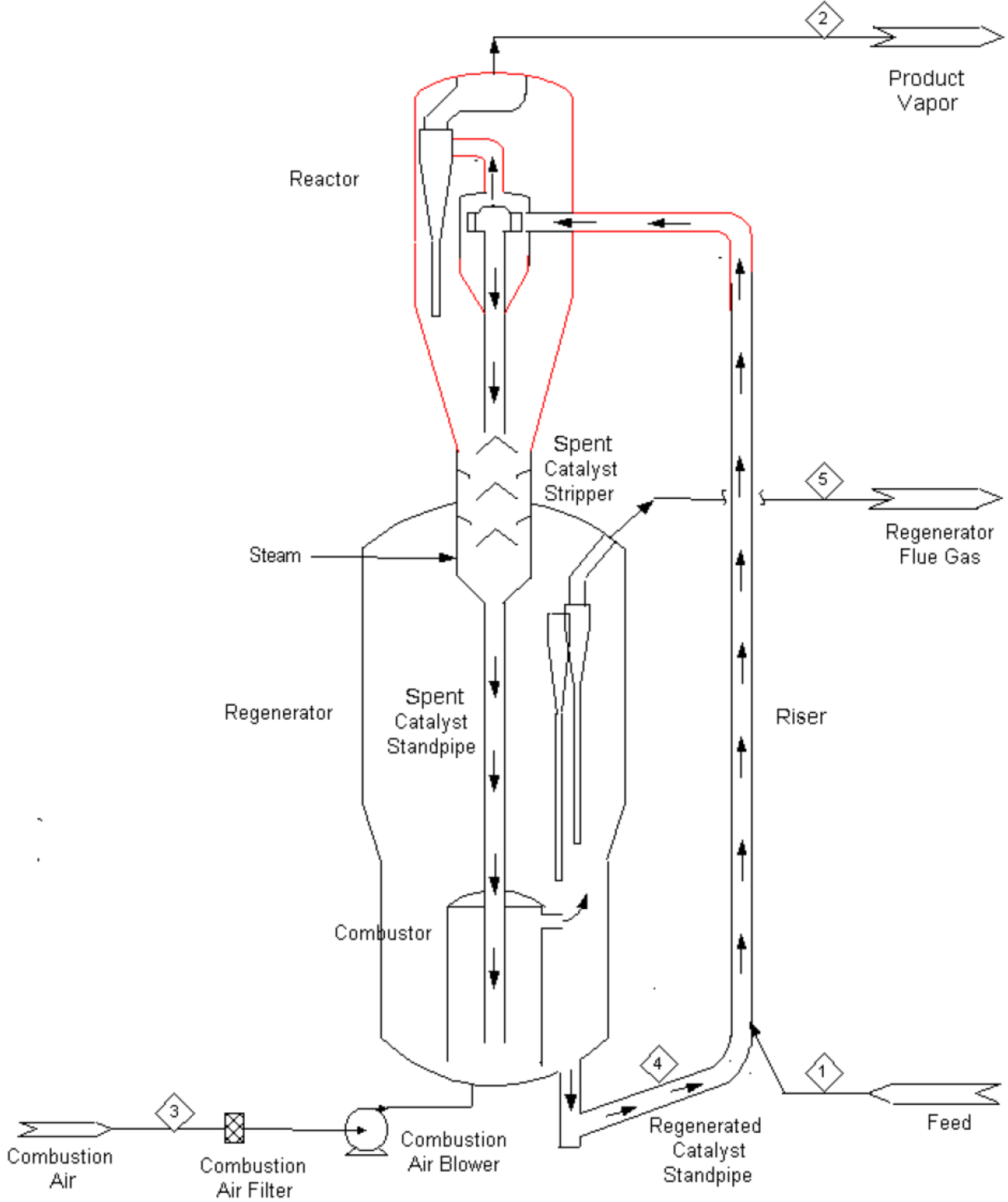

FIGURE 1-2 SMALL SCALE FCC DESIGN

TABLE 1-3 LARGE SCALE FCC EQUIPMENT LIST

						Design		Operating							
Item No	Description	Type	Quantity Per Train	Size	Weight	P	T	P	T	Power Useage	Materials	Price, total (unistalled)	Total Installed Cost	Quote Source	Comments
					lbs	PSIG	ºF	PSIG	ºF	(No.) HP		2002 Cost Index $US	(US $)		
Reactors															
	Reactor Vessel	Fluidized Bed	1	27'-10"ID x 25'-0"T/T	1,300,000	50	1100	35	1000		CS+1/16" CA (Shell), 4" refractory	$9,958,080		Major EPC Vendor	
	--Stripper (Part of Reactor)	Fluidized Bed	1	13'-7"ID (bottom of reactor)	incl. w/ reactor	50	1100	35	1000		304SS+1/16" CA, 4" refract.	incl. with reactor		Major EPC Vendor	Included with Reactor Cost and weight
	Regenerator Vessel	Fluidized Bed	1	43'-8"ID x 35'-0" T/T	incl. w/ reactor	50	1600	35	1300		CS+1/16" CA (Shell), 4" refractory	incl. with reactor		Major EPC Vendor	
	--Combustor (Part of Regenerator)	Fluidized Bed	1	38'-0"ID x 15'-0" T/T	incl. w/ reactor	50	1600	35	1300		304SS+1/16" CA, 4" refract.	incl. with reactor		Major EPC Vendor	Included with Reactor Cost and weight
Total												**$9,958,080**			
Cyclones															
	Reactor Cyclones	Cyclone	4	10 FT2 inlet, 11 FT2 outlet, 2.5' dipleg, 2 stage	42,400/unit	50	1100	35	1000		1-1/4 Cr -1/2 Mo + 1/8" CA	$910,200		Major EPC Vendor	
	Regenerator Cyclones	Cyclone	10	5.5 FT2 inlet, 6.3 FT2 outlet, 2' dipleg, 2 stage	35,000/unit	50	1600	35	1300		304SS + 1/16" CA	$2,484,600		Major EPC Vendor	
												$3,394,800			
Total															
Columns, Vessels & Tanks															
	Riser	Vertical	1	5'-9"ID x 98'-0" T/T	incl. w/ reactor	50	1600	35	1300		CS+1/16" CA, 4" refract.	incl. with reactor		Major EPC Vendor	Included with Reactor Cost and weight
	Fresh Catalyst Storage Drum	Vertical	1	15'-6"ID x 62'-0" T/T	112,000	50	650	50	650		CS+1/16" CA	$214,020		Major EPC Vendor	
	Spent Catalyst Storage Drum	Vertical	1	15'-6"ID x 62'-0" T/T	112,000	50	650	50	650		CS+1/16" CA	$214,020		Major EPC Vendor	
Total												**$428,040**			
Compressors & Blowers															
	Combustion Air Compressor	Centrifugal	1	171000 ACFM						20,300	CS	$8,085,000		Major EPC Vendor	
Total												**$8,085,000**			
Other															
	Reactor Vent Silencer	Vertical	1	4'-6"ID	incl. w/ reactor	50	1100	35	1000		CS+1/16" CA	$41,820		Major EPC Vendor	
Total												**$41,820**			
SUBTOTAL												**$21,907,740**			
BULK MATERIALS												**$28,480,062**		Nexant/Major EPC Vendor	130% of Equipment Cost (instruments, piping, electrical, concrete, paint, etc.)
TOTAL PROJECT COST												**$50,387,802**	**$102,615,654**		Factor includes installation and indirect costs

TABLE 1-4 SMALL SCALE FCC EQUIPMENT LIST

						Design		Operating							
Item No	Description	Type	Quantity Per Train	Size	Weight	P	T	P	T	Power Useage	Materials	Price, total (unistalled)	Total Installed Cost	Quote Source	Comments
					lbs	PSIG	ºF	PSIG	ºF	(No.) HP		2002 Cost Index $US	(US $)		
Reactors															
	Reactor Vessel	Fluidized Bed	1	7' ID x 33' T/T	100,000	50	1100	35	1000		CS+1/16" CA (Shell), 4" refractory	$1,550,000		UOP/Nexant/EPC Vendor Estimate	
	--Stripper (Part of Reactor)	Fluidized Bed	1	3'-6"ID (bottom of reactor)	incl. w/ reactor	50	1100	35	1000		304SS+1/16" CA, 4" refract.	incl. with reactor		Nexant Estimate	Included with Reactor Cost and weight
	Regenerator Vessel	Fluidized Bed	1	9"-6" ID x 40'-0" T/T	incl. w/ reactor	50	1600	35	1300		CS+1/16" CA (Shell), 4" refractory	incl. with reactor		UOP/Nexant/EPC Vendor Estimate	
	--Combustor (Part of Regenerator)	Fluidized Bed	1	8'-0"ID x 6'-6" T/T	incl. w/ reactor	50	1600	35	1300		304SS+1/16" CA, 4" refract.	incl. with reactor		Nexant Estimate	Included with Reactor Cost and weight
Total												**$1,550,000**			
Cyclones															
	Reactor Cyclones	Cyclone	2	2 foot diameter, single stage	3,100/unit	50	1100	35	1000		1-1/4 Cr -1/2 Mo + 1/8" CA	$30,000		Nexant/EPC Vendor Estimate	
	Regenerator Cyclones	Cyclone	3	3 foot diameter, two stage	5,200/unit	50	1600	35	1300		304SS + 1/16" CA	$90,000		Nexant/EPC Vendor Estimate	
												$120,000			
Total															
Columns, Vessels & Tanks															
	Riser	Vertical	1	2' ID x 85'-0" T/T	incl. w/ reactor	50	1600	35	1300		CS+1/16" CA, 4" refract.	incl. with reactor		Nexant Estimate	Included with Reactor Cost and weight
	Fresh Catalyst Storage Drum	Vertical	1	3'-6"ID x 12'-0" T/T	20,000	50	650	50	650		CS+1/16" CA	$36,900		Nexant Estimate	
	Spent Catalyst Storage Drum	Vertical	1	3'-6"ID x 12'-0" T/T	20,000	50	650	50	650		CS+1/16" CA	$36,900		Nexant Estimate	
Total												**$73,800**			
Compressors & Blowers															
	Combustion Air Compressor	Centrifugal	1	7000 ACFM						1,600	CS	$600,000		Nexant/EPC Vendor Estimate	
Total												**$600,000**			
Other															
	Reactor Vent Silencer	Vertical	1	1' ID	incl. w/ reactor	50	1100	35	1000		CS+1/16" CA	$6,765		Nexant Estimate	
Total												**$6,765**			
SUBTOTAL												**$2,350,565**			
BULK MATERIALS												**$3,055,735**		Nexant/EPC Vendor Estimate	130% of Equipment Cost (instruments, piping, electrical, concrete, paint, etc.)
TOTAL PROJECT COST												**$5,406,300**	**$11,010,025**		Factor includes installation and indirect costs

Two cost factors were used on top of the equipment cost to come up with the total project investment (TPI). First, the equipment cost was multiplied by 1.3 to estimate the cost for all bulk materials. This includes items such as instruments, piping, concrete, electrical, and all associated labor involved with fabricating these items. An installation and project cost factor of 2.04 was applied to the equipment cost to arrive at the TPI. This not only includes installation, but also all indirect costs. Both factors were based upon Nexant and vendor estimates for a detailed FCC design. The results of each case can be seen in Table 1-5.

TABLE 1-5 TOTAL FCC PROJECT COSTS, 2002 BASIS

	Large Scale Plant	Small Scale Plant
Feed Rate (BPSD)	62,000	2,500
Reactor/Regen TPI ($MM)	102.6	11.0
Total Project Investment ($MM)	191.6	20.5
Total Investment ($/bbl)	~3,000	~8,200

As can be seen from Table 1-5, there is a significant economy of scale advantage realized in the construction of larger FCC units. In fact, the $/bbl investment for even larger plants (~100,000 BPSD) may approach ~$2000[4]. This is a large reason why small FCC units are no longer built, except in specialty niche applications where petroleum supply and product demand are limited but stable. In general, petroleum companies are building all refinery process units large to improve refining margins.

1.3.1 Large Scale (Side-by-Side) FCC

Riser Section

In the "side-by-side" design, the riser is inside the steam stripper, as shown in Figure 1-1. The operating pressure ranges from 8 to 30 psig, and is generally limited to 15 to 20 psig by the design and pressure balance of the reactor and the regenerator. Lower operating pressure tends to improve yield selectivity, enhance stripper performance, and reduce air blower requirements, but increases vessel sizes and investment cost.

The riser temperature ranges from 950°F to 1025°F. It is controlled by the catalyst circulation rate, which in turn is controlled by the position of the slide valve at the bottom of the Regenerated Catalyst Standpipe. Increasing temperature, catalyst/oil ratio, catalyst activity, and contact time results in higher conversion and significant octane improvement. However, cracking at higher temperatures also produces more gas and coke which are less desirable products. The optimum temperature is when just enough coke is formed on the catalyst so that the heat of combustion in the regenerator can provide all the heat required for reaction when the catalyst returns to the riser-reactor.

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[4] Information from Mark Schnaith, UOP FCC Business Development Manager, personnel communications, and "Refinery Processes 2002", Hydrocarbon Processing, November 2002. The actual investment will depend on the technology chosen, the gas oil feed properties, and the products desired.

Since modern catalysts are highly active, conversion is controlled by varying catalyst circulation rate. A short residence time of the gas in the riser, usually less than 5 seconds, is maintained in order to produce the desired product selectivity and amount of coke formation.

Fresh catalyst is added to maintain the desired activity of the catalyst as the activity slowly declines from residual coke deposit and deterioration. Spent catalyst is removed periodically to maintain the inventory in the desired range.

Reactor Section

The reactor serves as a disengaging vessel for the catalyst and product vapor, and as the housing for the cyclones. Operating pressure and temperature of the reactor are similar to those at the riser.

At the top of the riser, all the selective cracking reactions are terminated by separation of the product vapor from the spent catalyst. Since over-cracking diminishes the value of the product, it is important to minimize the product vapor residence time in the reactor disengaging vessel. Special mechanical riser termination devices have been developed to ensure rapid separation of the vapor from the catalysts. These devices include vortex separation system developed by UOP, vented risers, and close-coupled cyclones often used by KBR/EOM. Cyclones mounted inside the reactor removes the catalysts from the product vapor before entering the fractionator and gas recovery sections. The diplegs of these cyclones are sealed in the fluidized catalyst bed or with counter weighted valves.

Spent catalyst separated from the cyclones contains entrained and adsorbed hydrocarbons as well as the coke deposited by the cracking reaction. Removing the hydrocarbons from the catalyst before regeneration minimizes the regenerator bed temperature, regeneration air requirement, and flue gas volume. Spent catalyst flows down into the Spent Catalyst Stripper which is connected to the base of the reactor vessel. In this section, ascending stripping steam is used to remove as many hydrocarbons as possible from the catalyst as it descends through the stripper. Hydrocarbon vapors and steam flow upwards into the reactor disengaging vessel and mix with the product vapors from the riser. The stripped spent catalyst descends through the Spent Catalyst Standpipe into the regenerator. A refractory-lined steel slide valve at the base of this standpipe provides control of the catalyst circulation rate.

In the “side-by-side” design, the reactor vessel is located next to the regenerator vessel. The Spent Catalyst Standpipe and the Regenerated Catalyst Standpipe connect these two vessels, as shown in Figure 1-1.

Regenerator Section

The regenerator is a fluidized bed where the activity of the spent catalyst is restored. Coke deposited on the catalyst during the cracking reactions is burnt off with air. Pressurized air is injected into the regenerator to provide oxygen for coke combustion and bed fluidization. The exothermic process of coke combustion heats the catalyst which transports the heat energy to the riser for the cracking reactions. Cyclones mounted inside the regenerator vessel remove the catalyst carryover from the flue gas before it is cooled and treated for release to the atmosphere.

Both small and large scale units have a combustor style regenerator. This design features a fast fluidized bed as the low inventory carbon burning zone followed by a higher velocity transport-riser heat exchange zone. It provides uniform coke-air distribution which enhances the ability to burn almost all the coke to carbon dioxide, leaving very low levels of carbon monoxide in the flue gas. The regenerated catalyst has low residual carbon and high effective activity, and is returned to the riser through the Spent Catalyst Standpipe.

Typical regenerator operating pressure ranges from 15 to 30 psig. The regenerators are generally operated at the maximum practical temperature from 1,200 to 1,500^{o}F for maximum combustion of the coke deposits and preservation of catalyst activity and selectivity. The higher the differential temperature between the regenerator and the reactor, the lower the catalyst circulation needed.

Design Assumptions

Since the methodology for design of an FCC of this scale is well-established, a high level of confidence exists in the estimates presented here. Although this design was originally performed in the early 1990's, the major design aspects have not changed. Few assumptions were required due to the quality of the existing data. Assumptions used in the design include:

- Cost Escalation: Factors were used to escalate the early 1990's cost data to 2002. These factors largely came from the ICARUS database. Not all items escalated at the same rate; for example, the escalation on compressor costs was considerably less than on the reactor/regenerator vessels.
- Scope: Per NREL guidelines, an entire FCC design is not presented here. The items deemed to be most important for future NREL use are the reactor/regenerator and associated equipment. Since all vendor supplied FCC design packages are for an entire FCC facility including downstream fractionation and product treating, splitting the design to just focus on the reactor/regenerator made it difficult to estimate utility requirements. Main process utilities were therefore estimated from literature estimates instead of vendor packages.
- Internals: Items such as the combustor, riser, cyclones, and stripper are usually provided as a package in the FCC design estimate, with limited design information provided. Some of these pieces of equipment are considered proprietary by vendors.

1.3.2 Small Scale (Stacked) FCC

In general, all the major licensors agree that the "stacked" design requires less plot space and less structural steel, thus lowering the installed cost. As capacity increases, the vessel sizes also increase to a point where structural cost exceeds this advantage. As a rule of thumb, a capacity of 20,000 BPSD is the cut-off size where "side-by-side" design becomes more economical than "stacked" design, according to UOP. There are other factors that affect the cost/BPSD for the same capacity and thus affect the selection of the FCC configuration. These factors include location, available land space, and local building codes.

In general, the theory behind the design of the small-scale unit is very similar to that of the large-scale unit. Regardless of the variation of design by the licensors, the components that make up the FCC are the same: the riser, the reactor, and the regenerator. Two design differences can be seen in the riser and reactor for the stacked design:

- The riser is outside the stack of reactor, stripper and regenerator vessels. The product vapors and catalyst flow from the riser into the reactor vessel through a nozzle.
- The reactor vessel is mounted on top of the regenerator vessel. The Spent Catalyst Standpipe is located inside the regenerator vessel. The Regenerated Catalyst Standpipe connects the regenerator to the external riser.

Design Assumptions

Less vendor detail and previous design information was available for estimation of the small scale FCC. In general, small scale FCCs are no longer economic, and very few units under 20,000 BPSD are being built today. UOP was able to provide some design information for its 2,500 BPSD unit in Great Falls, Montana, which was built in the 1950's. A greater level of assumptions and scaling was required when developing this design estimate:

- Reactor/Regenerator: UOP provided the sizes of both the reactor and regenerator, but not the cost. The cost was developed through the use of ICARUS and Nexant design information. Internals were estimated from literature sources.
- Cyclones: The quantity, dimensions, and cost for the cyclones were developed through the use of ICARUS and vendor input.
- Air Compressor: Literature factors relating the amount of process air to the feedrate was used as the basis for the air compressor sizing. Horsepower requirements and costs were obtained through the use of HYSYS and ICARUS.

1.4 OPERATING COSTS AND UTILITY REQUIREMENTS

The basis for the operating costs and utility requirements are shown in Table 1-6.

TABLE 1-6 FCC OPERATING COSTS AND UTILITY REQUIREMENTS

Operating Variable	Large Scale	Small Scale	Notes
Catalyst[5] *Initial Load* *Make-Up*	 468 tons 6 tons/day	 19.5 tons 0.25 tons/day	$2000/ton catalyst cost[6]. Assumes no enhanced olefins production
Combustion Air	146 MMSCFD	6 MMSCFD	Assumes no excess air
Medium Pressure Steam Consumption (React/Regen Only)[7]	68,333 lb/hr	1,424 lb/hr	Side-by-side plant requires additional steam
High Pressure Steam Production (Overall FCC)[8]	100,000 lb/hr	4,167 lb/hr	Overall FCC is net steam long
Electricity Use[9]	15 MW	0.6 MW	Assumes electric driver for air compressor

The majority of the estimates used here come from literature sources. When sources differed in their estimates, an average was assumed to develop a “typical” FCC operating cost. For the most part, FCC utility requirements can be fairly accurately predicted just by knowing the unit feed rate: catalyst, combustion air, electricity, and steam requirements can all be roughly estimated by knowing this key quantity. A few key items for consideration when developing overall operating costs:

- Daily catalyst make-up requirements can vary considerably based on the age of the unit, the outlet composition desired, and the efficiency of the cyclones.
- Most FCC units use some excess process air to assure complete combustion of the coke deposited on the catalyst. Since this can vary considerably, a base design assumption without process air is used for both units. The design for the air compressors assumes capacity available for excess process air.
- Steam costs are difficult to estimate, for an FCC will be integrated into an overall refinery configuration. It is likely that sufficient process steam will be available to keep the cost of this utility low.

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[5] Data compiled from the following sources: “Refinery Processes 2002”, Hydrocarbon Processing, November 2002, “Petroleum Refining—Technology and Economics” by J. H. Gary and G. E. Handwerk, 1984, and The FCC Network, Volume 4, June-July 2000, available at http://thefccnetwork.com/pdf/newsletters/newsletter4.pdf

[6] Information from Mark Schnaith, UOP FCC Business Development Manager, personnel communications, 15 August 2005.

[7] Data gathered from Bechtel survey of FCC utility requirements, 1981. Average of UOP designs taken, with distinctions made between stacked and side-by-side design requirements.

[8] Information from KBR data in Hydrocarbon Processing, November 2002, and UOP information from Meyers R., “Handbook of Petroleum Refining Processes”, Third Edition, 2004.

[9] Bechtel survey of UOP equipment, 1981. Data checked versus expected compressor horsepower requirements.

2.1 INTRODUCTION AND METHODOLOGY

Detailed designs for both large and small scale Steam Methane Reforming (SMR) units were developed specifically for hydrogen production. The small scale design assumes a methane feed rate of 0.024 MMSCFD (producing 10 lb/hr of hydrogen), while the large scale design uses a methane feed rate of 168 MMSCFD (producing ~100,000 lb/hr of hydrogen). These feed rates represent a typical size for two distinct types of SMR facilities: the small scale plant is consistent with hydrogen production for small fuel cell applications, while the large facility would be seen at a petrochemical or ammonia plant.

Data was collected from publications and licensors of SMR units. The focus of this research was on the capital and operating costs for entire large and small scale units including natural gas purification, reforming, shift, and hydrogen purification. The compression and storage of the hydrogen product are not included. Design and cost information was obtained primarily from literature sources and process modeling. A number of sources, noted in this report, were used for estimating capital and operating costs.

2.2 TECHNOLOGY BACKGROUND

Natural gas reforming is the process of converting natural gas feedstocks to a synthesis gas consisting of CO, CO_2, and H_2 using steam as a reactant. The synthesis gas will also contain small amounts of unreacted methane and residual steam. This process is the first step in many industrial processes including methanol synthesis, ammonia manufacture, gas-to-liquids (GTL), and hydrogen production. The earliest SMR units were introduced in the U.S. in the 1930s by Standard Oil of New Jersey at their Bayway and Baton Rouge refineries for hydrogen production from offgases[10].

Steam methane reforming is the most common and least expensive route to producing hydrogen. The technology is well developed and commercially available at a wide capacity range, from less than 1 MMSCFD of methane feed for fuel cells, to 400 MMSCFD of methane feed for ammonia production. The process involves the catalytic reaction between methane and steam to produce a synthesis gas. The synthesis gas is further processed through another catalytic step to increase the hydrogen fraction and decrease the carbon monoxide via the water gas shift reaction. Finally, a purification step removes all other components to produce the hydrogen product.

2.3 PROCESS DESCRIPTIONS AND DESIGN ASSUMPTIONS

Similar to the information presented in the FCC section, flow diagrams, stream properties, and an equipment list for both cases is presented below. Due to the larger overall process scope and greater process complexity, detailed stream properties are shown for each case.

Unlike the FCC design case, unique installation factors specific to SMR units have not been used or identified by the Nexant team. Therefore, a process similar to that used by NREL in the recent “Biomass Syngas to Hydrogen Production Design Report” was used. After calculation of

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[10] Catalyst Handbook, Second Edition, Martyn V. Twigg, ed., Wolfe Publishing Ltd., 1989

the equipment cost, a factor of 2.47 was used to calculate the Total Installed Cost (TIC). For indirect costs, factors developed by NREL to account for engineering costs, construction, contingencies, and fees were applied. Indirect cost factors were then added to the TIC to obtain the Total Project Investment (TPI). A summary of the results for both cases can be seen in Table 2-1.

TABLE 2-1 SMR PROJECT COSTS, 2005 BASIS

	Large Scale Plant	Small Scale Plant
Natural Gas Feed Rate (MMSCFD)	167.8	0.024
Hydrogen Production Rate (lb/hr)	101,885	10.6
Equipment Cost ($MM)	98.8	0.10
Installed Cost ($MM)	244.0	0.26
Total Plant Investment ($MM)	368.5	0.39
Total Investment/SCFD Feed ($)	2.2	16.2

Due to the very different plant sizes being considered, a large economy of scale difference is evident. Roughly 7 times the investment per SCFD of methane feed is required to produce small amounts of hydrogen relative to the largest plants in operation today. While the capital required is significant to capture this economy of scale, large scale centralization of hydrogen production appears to have a significant cost advantage if large amounts of hydrogen is desired.

2.3.1 Small Scale SMR Design

Figure 2-1 shows the PFD for the small scale SMR unit, while Table 2-2 shows the stream conditions. The equipment list for the small scale design can be found in Table 2-3.

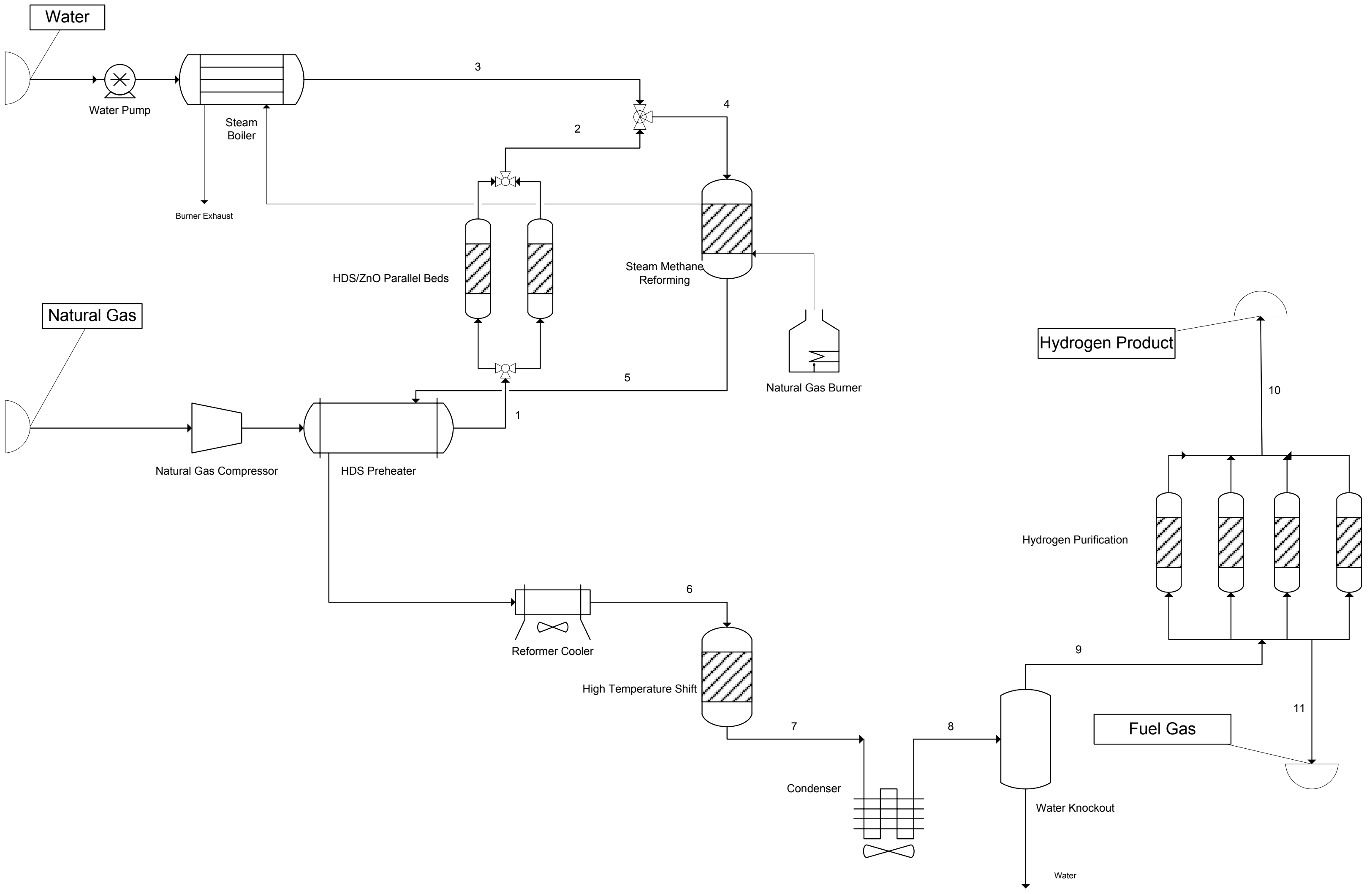

FIGURE 2-1 SMALL-SCALE SMR PROCESS FLOW DIAGRAM

TABLE 2-2 SMALL SCALE SMR STREAM PROPERTIES

Stream ID	Natural Gas Feed	Desulfurized Natural Gas	Reaction Steam	Reformer Feed	Reformed Product
Stream Number	1	2	3	4	5
Methane (lb-mole/h)	2.13	2.13		2.13	0.63
H_2O	0.00	0.00	7.36	7.36	4.76
CO					1.01
CO_2	0.02	0.02		0.02	0.82
Hydrogen					6.04
Ethane	0.08	0.08		0.08	0.00
Propane	0.04	0.04		0.04	0.00
Oxygen	0.02	0.02		0.02	0.02
Nitrogen	0.04	0.04		0.04	0.04
Total lbmole/h	2.4	2.4	7.4	9.7	13.3
Mass Flow [lb/hr]	41.7	41.7	132.6	174.3	174.3
Std Gas Flow [cu ft/hr]	892	892	2,793	3,685	5,056
Std Gas Flow [MMSCFD]	0.02	0.02	0.07	0.09	0.12
Temperature [F]	850	850	850	847	1472
Pressure [psia]	295	295	295	295	295

TABLE 2-2 SMALL SCALE SMR STREAM PROPERTIES (Cont'd)

Stream ID	Shift Feed	Shifted Gas	Cooled Shifted Gas	Shift Condensate	PSA Feed	Hydrogen Product	PSA Purge
Stream Number	6	7	8		9	10	11
Methane (lb-mole/h)	0.63	0.63	0.63	0.00	0.63	0.00	0.63
H_2O	4.76	3.79	3.79	3.70	0.08	0.00	0.08
CO	1.01	0.04	0.04	0.00	0.04	0.00	0.04
CO_2	0.82	1.79	1.79	0.00	1.79	0.00	1.79
Hydrogen	6.04	7.01	7.01	0.00	7.01	5.26	1.75
Ethane	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Propane							
Oxygen	0.02	0.02	0.02	0.00	0.02	0.00	0.02
Nitrogen	0.04	0.04	0.04	0.00	0.04	0.00	0.04
Total lbmole/h	13.3	13.3	13.32	3.7	9.6	5.3	4.4
Mass Flow [lb/hr]	174.3	174.3	174.3	66.8	107.4	10.6	96.8
Std Gas Flow [cu ft/hr]	5,056	5,056	5,056	1,406	3,650	1,996	1,654
Std Gas Flow [MMSCFD]	0.12	0.12	0.12	0.03	0.09	0.05	0.04
Temperature [F]	840	480	130	130	130	130	130
Pressure [psia]	290	290	285	280	275	420	3

TABLE 2-3 SMALL-SCALE SMR EQUIPMENT LIST

						Design		Operating							
Item No	**Description**	**Type**	**Quantity Per Train**	**Size**	**Weight**	**P**	**T**	**P**	**T**	**Power Useage**	**Materials**	**Price, total (unistalled)**	**Total Installed Cost**	**Quote Source**	**Comments**
					lbs	PSIG	ºF	PSIG	ºF	HP		2002 Cost Index $US	(US $)		
Reactors															
	Hydro-Desulfurization Bed	Fixed Bed	2	4" pipe, 30" length		350	1,000	295	850		316 SS	$5,277		Directed Technologies Design	
	Reforming Bed	Fixed Bed	1	Shell: 14" D, 40" L Tubes: 134 @ 1/2" D, 39" L		350	1,600	295	1,475		Incoloy 800H (Shell), Haynes 556 (Tubes)	$28,746		Directed Technologies Design	Includes natural gas burner
	High Temperature Shift	Fixed Bed	1	Shell: 15.25" D, 75" L Tubes: 106 @ 1" D, 75" L		350	1,000	295	840		316 SS	$20,594		Directed Technologies Design	Includes reformer cooler
	Pressure Swing Adsorber	Fixed Bed	4	12" D, 72" L		350	150	295	130		316 SS	$18,075			
Total												**$72,692**			
Exchangers															
	HDS Preheater	Shell/Tube	1	Shell: 2.5" Sched 40, 12" long Tube: 1" D, 23" L (1 tube)		350	1,000	295	850		316 SS, Incoloy	Included with HDS Bed		Directed Technologies Design	HDS Preheater and Reformer Cooler Combined Duty of 0.0746 MMBTU/hr
	Steam Boiler	Shell/Tube	1	Shell: 16.5" D, 320" L Tubes: 64 1" tubes		350	1,000	295	850		316 SS (shell), Incoloy (tubes)	$18,415		Directed Technologies Design	
	Reformer Cooler	Air Fan	1	13 Tubes, 10"D, 18" L		350	1,600	295	1,475		316 SS (shell), Haynes 556 (tubes)	Included with HT Shift reactor		Directed Technologies Design	HDS Preheater and Reformer Cooler Combined Duty of 0.0746 MMBTU/hr
	Post-Shift Condenser	Air Fan	1	3.5 cu. ft, stainless steel finned tubes		350	600	295	480		316 SS finned tubes	$4,729		Directed Technologies Design	Duty = .104 MMBTU/hr
Total												**$23,144**			
Pumps, Compressors & Blowers															
	Natural Gas Compressor	Centrifugal	1	15 ACFM		350	500	295	200	8.5 HP	CS	$3,354		GTI	
	Water Pump	Pos. Displacement	1	1.5 L/min, 300 psig		350	200	200	300	0.06 HP	CS	$3,210		GTI	
Total												**$6,564**			
SUBTOTAL												**$102,400**			
TOTAL PROJECT COST													**$252,928**		

Natural gas (stream 1) is compressed from near atmospheric delivery pressure to 295 psia to decrease downstream equipment sizes and take advantage of the increased reaction rates and driving forces seen at higher pressures. The gas is then heated in the hydrodesulfurization (HDS) preheater to 750°F before entering the HDS/ZnO desulfurization beds. This step is necessary because the reforming catalysts are poisoned by even trace quantities of hydrogen sulfide, mercaptans, and odorants. The HDS catalyst converts mercaptans to hydrogen sulfide while the zinc oxide adsorbs the hydrogen sulfide. Two parallel beds are used to allow undisturbed operations. Once the bed material is fully loaded, the bed can be emptied and refilled while the other bed is treating the gas.

Steam (stream 3) is produced by vaporizing water in a boiler heated by a natural gas burner. Steam is necessary for heat during the endothermic reforming reaction and for use in the water-gas shift step. The steam and hot desulfurized methane stream (stream 4) are combined and fed to the steam methane reformer. This reactor contains a nickel-based catalyst where the following reactions take place:

$$CH_4 + H_2O \Leftrightarrow CO + 3H_2 \text{ (reforming)}$$

$$CO + H_2O \Leftrightarrow CO_2 + H_2 \text{ (water-gas shift)}$$

The reforming reaction is highly endothermic, with the heat of reaction supplied by an external fired heater. The steam to carbon ratio typically varies from 2.5 to 5 on a molar basis. The equilibrium conversion of methane at 295 psig is 73%. A depiction of a typical small scale SMR reactor can be seen in Figure 2-2. Designs follow fairly closely with standard heat exchanger design to minimize unique requirements that lead to increased cost.

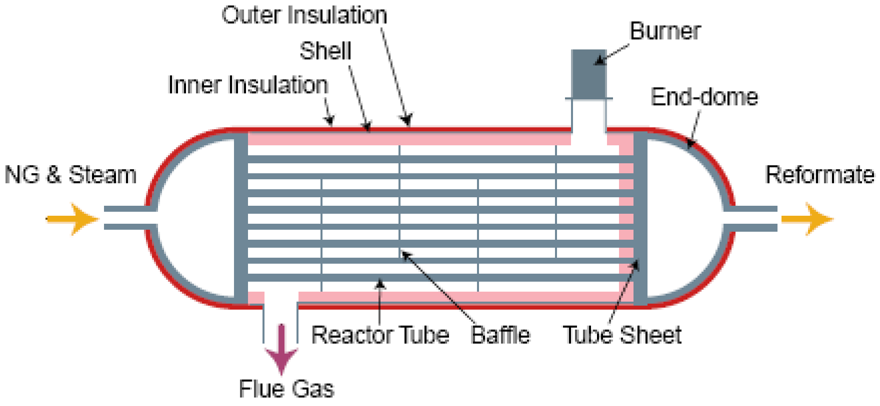

FIGURE 2-2 TYPICAL SMALL SCALE SMR REACTOR DESIGN[11]

The resulting synthesis gas is cooled through the HDS preheater and reformer cooler to enter the high temperature shift reactor (stream 6). Here the water gas shift reaction occurs over a promoted iron oxide catalyst to increase the amount of hydrogen while decreasing the carbon monoxide content. Over 95% conversion of the carbon monoxide is achieved.

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[11] "Cost and Performance Comparison of Stationary Hydrogen Fueling Appliances, Task 2 Report" Myers, D.B., et al., Directed Technologies, Inc. (April 2002)

The next step is to cool the gas stream in order to condense the water vapor in the water knockout vessel. Pressure swing adsorption (PSA) is then used to produce the final hydrogen stream (stream 10) at 99.99% purity and with less than 10 ppm carbon monoxide. The remaining hydrogen, methane, carbon dioxide, carbon monoxide, nitrogen, and other trace components are recovered in the regeneration step and used as a fuel gas (stream 11). The PSA system uses a number of adsorbent beds, containing a 5Å molecular sieve to selectively pass hydrogen while adsorbing the other components. Once loaded with these components and before they breakthrough the bed into the hydrogen product, the bed is taken off-line for regeneration while a parallel bed is put on-line with no break in the continuity of gas flow. The pressure in the regenerating bed is dropped to drive off the adsorbed components. The cycle is run automatically on a timed schedule to avoid breakthrough. A four-bed system is used in this design.

Design Assumptions

Since this study includes both large and small scale designs, two PFDs have been developed to reflect the type of configuration which is considered most common and economic for the unit size. This has led to two different designs:

- A reformer that integrates the external heat to drive the endothermic reforming reactions and convection heat exchange to provide heating and cooling of the process streams. This design is typically used for medium and large systems, with the basis used for the large scale unit design.
- A stand-alone reformer with external heating provided by natural gas and non-integrated heat exchange systems designed to minimize cost. This was used for the small scale unit. The type of heat integration and reformer design used in the large scale unit is not practical for hydrogen production at the small scale. Creating designs that exactly replicated the technology used in either case would not be indicative of true commercial applications. For example, air coolers for the reformer cooler and post-HTS condenser are considerably more practical in the small scale due to the low duty required.

The small scale plant assumes that the design can be used anywhere hydrogen is required, and does not require integration into an industrial facility. Compressors and pumps are included for natural gas and water, with the assumption that neither of these feeds is available at the conditions required in the process. Fuel gas produced in the process can be used in the natural gas burner to help provide the reformer heat duty.

2.3.2 Large Scale SMR Design

Figure 2-3 shows the PFD for the large scale SMR unit, while Table 2-4 shows the stream conditions. The equipment list for the large scale design can be found in Table 2-5.

The major difference between the small and large scale systems is the addition of an internal direct-fired heater to provide the necessary heat of reaction for the large volume of natural gas feed. At this size, the reformer reactor is usually integrated with the fired heater assembly in a packaged system. In addition, the large scale system uses additional heat recovery systems to utilize heat from the direct fired heaters. The heat integration increases the thermal efficiency of the plant relative to the small scale unit.

The natural gas feed stream is heated in the reformer exhaust system (stream 1) and passes through the desulfurization vessels where H_2S is adsorbed on a solid adsorbent. At least two beds are operated in parallel so that one unit can be recharged when the bed material is fully loaded without shutting down flow.

The desulfurized gas (stream 2) is mixed with steam (stream 3) and superheated in the feed preheat coil. The feed mixture (stream 4) then passes through reforming catalyst-filled tubes in the reformer vessel. In the presence of the nickel-based catalyst, the hydrocarbons (mainly methane) react with the steam to produce a synthesis gas composed mainly of carbon monoxide and hydrogen. The endothermic heat of reaction is provided by controlled external firing in the reformer. The exhaust gas from the fired heater provides a bank of convection coil heating for:

- Steam generation
- Feed/steam preheating
- Steam superheating
- Feed preheating
- Condensate preheating, and
- Combustion air preheating

The synthesis gas (stream 5) leaving the reformer is cooled by the process gas cooler/steam generator. Gas is then fed to the high temperature shift reactor (stream 6). This catalyst is copper promoted iron-chromium, which converts carbon monoxide and water to additional hydrogen with a coproduct of carbon dioxide. The shift reactor effluent (stream 7) is cooled in the condensed water heater, a boiler feed water (BFW) preheater, and a gas trim cooler. The cooled stream (stream 8) flows to a condensate separator where the remaining condensate is separated and the gas (stream 9) is sent to a PSA Hydrogen Purification System.

The PSA system is on an automatic cycle to minimize operational attention. It cycles several vessels between adsorption and regeneration cycles where (1) essentially all the gas components except hydrogen are adsorbed on a molecular sieve material, allowing the hydrogen to pass at pressure (stream 10), and then (2) desorption of the adsorbed components by reducing the pressure in the vessel. The PSA off-gas (stream 11) is sent to the reformer to provide some of the fuel required for reforming.

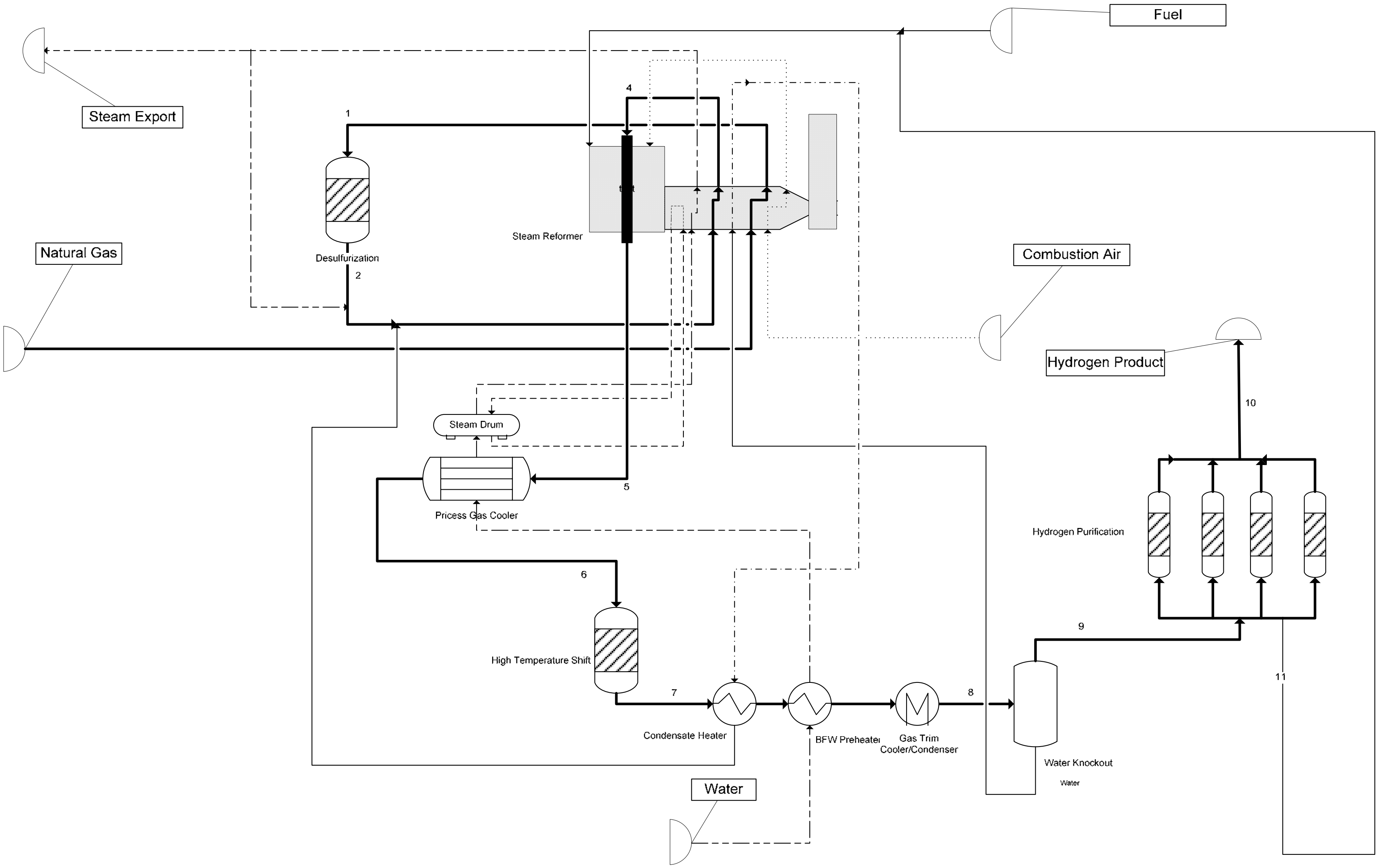

FIGURE 2-3 LARGE SCALE SMR PROCESS FLOW DIAGRAM

TABLE 2-4 LARGE SCALE SMR STREAM PROPERTIES

Stream ID	Natural Gas Feed	Desulfurized Natural Gas	Reaction Steam	Reformer Feed	Reformed Product
Stream Number	1	2	3	4	5
Methane (lb-mole/h)	16,707.4	16,707.4	-	16,707.4	3,864.7
H2O	2.1	2.1	49,166.0	49,168.1	29,142.8
CO	-	-	-	-	10,413.9
CO2	184.0	184.0	-	184.0	4,989.7
Hydrogen	-	-	-	-	49,101.1
Ethane	663.3	663.3	-	663.3	0.0
Propane	350.1	350.1	-	350.1	0.0
Oxygen	184.0	184.0	-	184.0	184.0
Nitrogen	331.1	331.1	-	331.1	331.1
Total lbmole/h	18,429.4	18,422.0	49,166.0	67,588.0	98,027.2
Mass Flow [lb/hr]	326,967	326,716	885,730	1,212,447	1,212,459
Std Gas Flow [MMSCF/hr]	7.0	7.0	18.7	25.6	37.2
Std Gas Flow [MMSCFD]	167.8	167.8	447.8	615.6	892.8
Temperature [F]	1000	1000	1000	996	1615
Pressure [psia]	430	430	430	430	430

Stream ID	Shift Feed	Shifted Gas	Cooled Shifted Gas	PSA Feed	Hydrogen Product	PSA Purge
Stream Number	6	7	8	9	10	11
Methane (lb-mole/h)	3,864.7	3,864.7	3,864.7	3,864.7	0.4	3,864.3
H2O	29,142.8	19,520.7	19,520.7	492.1	0.0	492.0
CO	10,413.9	791.8	791.8	791.8	0.1	791.7
CO2	4,989.7	14,611.7	14,611.7	14,585.8	1.5	14,584.3
Hydrogen	49,101.1	58,723.1	58,723.1	58,722.1	50,501.0	8,221.1
Ethane	0.0	0.0	0.0	0.0	0.0	0.0
Propane	0.0	-	-	-	-	-
Oxygen	184.0	184.0	184.0	184.0	0.0	183.9
Nitrogen	331.1	331.1	331.1	331.1	0.0	331.1
Total lbmole/h	98,027.2	98,027.2	98,027.2	78,971.5	50,503.0	28,468.5
Mass Flow [lb/hr]	1,212,459	1,212,456	1,212,456	868,507	101,885	766,622
Std Gas Flow [MMSCF/hr]	37.2	37.2	37.2	30.0	19.2	10.8
Std Gas Flow [MMSCFD]	892.8	892.8	892.8	719.2	460.0	259.3
Temperature [F]	840	525	130	130	130	130
Pressure [psia]	425	425	420	420	420	3

TABLE 2-5 LARGE-SCALE SMR EQUIPMENT LIST

						Design		Operating							
Item No	**Description**	**Type**	**Quantity Per Train**	**Size**	**Weight**	**P**	**T**	**P**	**T**	**Power Useage**	**Materials**	**Price, total (unistalled)**	**Total Installed Cost**	**Quote Source**	**Comments**
					lbs	PSIG	ºF	PSIG	ºF	(No.) HP		2002 Cost Index $US	(US $)		
Reactors															
	Hydro-Desulfurization Bed	Fixed Bed	8	16' 6" L, 8' 6" D	347,684	600	1000	430	750		SS316			GTI Design	4719 cu ft
	Reforming Bed	Fixed Bed	1	26' 0" L, 13' 6" D 1788 Tubes	546,360	600	1800	430	1650		Incoloy 800H (Shell), Haynes 556 (Tubes)			Uhde Design	7301 cu ft
	High Temperature Shift	Fixed Bed	2	27' 0" L, 14' 0" D	620,863	600	1000	430	840		SS316			Uhde Design	9983 cu ft
	Pressure Swing Adsorber	Fixed Bed	12	25' 0" L, 10' 0" D	161,508	600	150	430	130		SS316			Uhde Design	
Total												**$75,000,850**			Estimate; no breakdown provided by Udhe
Exchangers															
	Gas Trim Condenser	Shell/Tube	1	50 MMBTU/hr Duty		600	600	430	525		SS316			Uhde Design	Total duty with Condensate Heater is 446.2 MMBTU/hr; this split is an estimate
	Condensate Heater	Shell/Tube	1	396.2 MMBTU/hr Duty		600	600	430	525		SS316			Uhde Design	
	Process Gas Cooler	Shell/Tube	1	659.9 MMBTU/hr Duty		600	1000	430	1000		SS316 (shell), Haynes 556 (tubes)			Uhde Design	
	BFW Preheater	Shell/Tube	1	306.5 MMBTU/hr Duty		600	600	430	525		SS316			Uhde Design	
Total												**$23,214,575**			Estimate; no breakdown provided by Udhe
Pumps, Compressors & Blowers															
	Natural Gas Compressor														Not included--assumed supply at pressure from large scale case.
	Water Pump	Pos. Displacement	3	1839 GPM		600	100	430	70	683 HP	CS			GTI/Nexant	
Total												**$570,000**			Estimate; no breakdown provided by Udhe
SUBTOTAL												**$98,785,425**			
TOTAL PROJECT COST													**$244,000,000**		Uhde Estimate; no unit breakdown specified

Depictions of typical large scale SMR reactors and how they are integrated with fired heaters have been provided by Udhe and can be seen in Figures 2-4 and 2-5[12]. These designs are more unique than what is used in small scale designs. While this requires additional capital for plant construction, the net result is a more efficient facility.

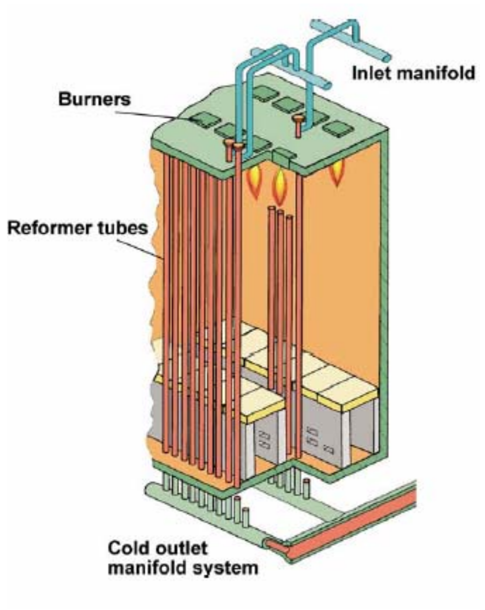

FIGURE 2-4 TYPICAL LARGE SCALE SMR REACTOR DESIGN

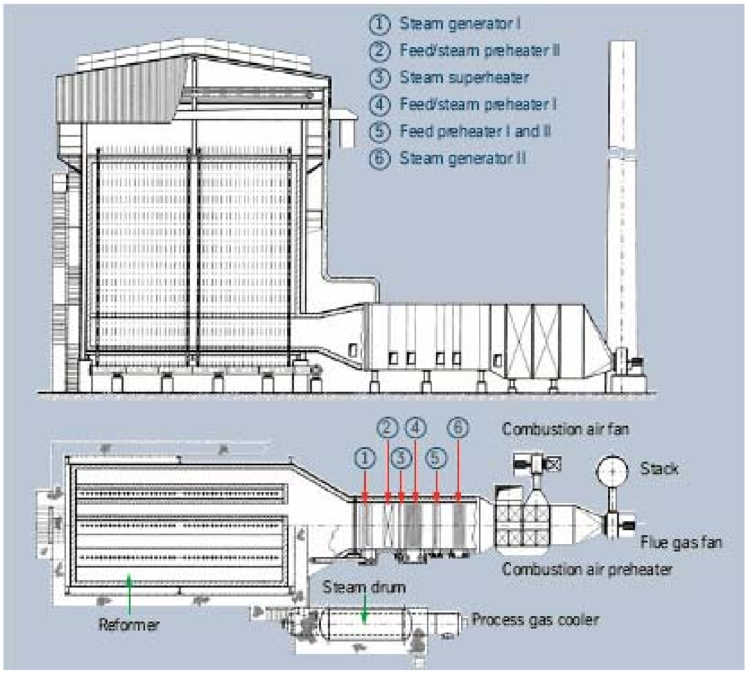

FIGURE 2-5 LARGE SCALE SMR REACTOR HEAT INTEGRATION

A small scale SMR will typically purchase natural gas from a utility at a pressure slightly above atmospheric pressure. The large systems will purchase their natural gas from a pipeline transmission company or directly from the producer, using the transmission company as a common carrier. In these systems, the pipeline pressures are 200 to 1500 psig. Therefore, a natural gas compressor is not included in this design. While a water pump is included in the large scale equipment list, a similar situation may occur where industrial water is available at pressure, making a pump unnecessary.

Table 2-6 highlights the main operating conditions for both the large and small scale cases. In general, the economy of scale in the large scale case allows more severe conditions, increasing the overall methane conversion and hydrogen recovery.

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[12] Hydrogen, UHDE, ThyssenKrupp, http://www.uhde.biz/cgi-bin/byteserver.pl/pdf/broschueren/Duengemittel/Hydrogen.pdf (August 2005)

TABLE 2-6 MAJOR PROCESS CONDITIONS, LARGE AND SMALL SMR DESIGNS

Process Conditions	Small Scale Design	Large Scale Design
Reforming pressure, atm	20	30
Steam to carbon ratio	3	2.8
Reformer inlet temp., °F	850	1000
Reformer outlet temp., °F	1470	1615
Methane conversion, %	71	77
Shift inlet temp., °F	840	840
Shift outlet temp., °F	480	525
CO concentration in shifted gas, %	0.3	0.8
PSA hydrogen recovery, %	75	86
Natural gas feed/hydrogen product, vol. ratio	0.47	0.36
Steam/hydrogen product, vol. ratio	1.4	0.97

2.4 OPERATING COSTS AND UTILITY REQUIREMENTS

The basis for the operating costs and utility requirements are shown in Table 2-7.

TABLE 2-7 SMR OPERATING COSTS AND UTILITY REQUIREMENTS

Operating Variable	Large Scale	Small Scale
Cooling Water (GPM)	1839	N/A
Natural Gas (MMSCFD)	15	0.002
Catalyst		
HDS/ZnO	1,415 ft^3/3,304 ft^3	0.06 ft3/0.14 ft^3
Reforming	7,300 ft^3	0.38 ft^3
Shift	9,983 ft^3	0.5 ft^3
PSA Adsorbent	250,000 ft^3	21 ft^3
Electricity Use (kW)	9517	16

The greater heat integration used in the large scale case leads to different utility requirements. For example, the small scale case uses mostly air-cooled exchangers, while the large scale case uses cooling water that is turned into steam for feed to the reformer reactor. Natural gas usage is predominantly for heat requirements in the reformer. Total electricity use comprises all power necessary for fans, blowers, motors, and starters.

Similar catalyst are used in both cases, with Co/W HDS catalyst, Fe/CrO shift catalyst, and, typically, reforming catalyst that is Ru/Ni/AlO. Catalyst costs and lifespans will likely be the same in both cases.

3.1 INTRODUCTION AND METHODOLOGY

Detailed large and small scale Natural Gas Liquid (NGL) Expander designs were developed assuming a super-rich (<85% methane) NGL feed. Providing detail for a super-rich design demonstrates to the full extent the amount of equipment necessary to perform an extensive fractionation on a multi-component system. As the feedstock becomes leaner, both the size of the plant and the amount of equipment necessary for separation decreases. The small scale design assumes 75 MMSCFD of inlet gas, while the large scale design has 400 MMSCFD of inlet gas. These are typical unit sizes for the small and large ranges of the NGL expander design spectrum.

The amount of ethane that is left in the methane stream depends largely on pipeline requirements and product economics. Current plants are designed to swing between ethane rejection (removed ethane is returned to the residual or product gas) and ethane recovery (removed ethane is included with the NGL product) depending on the operating conditions and how flows are directed in the plant. The small scale unit described below was based on ethane rejection, while the large scale unit recovers ethane. Regardless of the operating mode, the plant design will not change[13].

Data was collected from publications and licensors of NGL technology. The focus of this research was on the capital and operating costs of large and small scale designs based on turbo-expander technology. Since this study includes both large and small scale designs, configurations have been developed to reflect the design which is considered most common and economic for the unit size. The major difference between the designs is in the heat integration, with the larger-scale system utilizing more cross exchange of the hot and cold streams to improve overall heat efficiency. The large scale unit has also been modeled in HYSYS to confirm vendor information.

3.2 TECHNOLOGY BACKGROUND

Natural gas that is extracted from terrestrial sources consists predominantly of methane, and, to a lesser extent, a quantity of natural gas liquids. These are typically ethane and propane, but also contain some butane and lesser amounts of higher carbon content, aliphatic hydrocarbons. Table 3-1 below shows compositions for a range of potential feedstocks to a NGL plant.

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[13] A number of literature sources confirm this approach, including “Liquid Levels and Density Case histories with DP cell problems”, *Hydrocarbon Online,* March 2001, and “Deepwater Production Drives Design of New Gulf Gas Plant”, *Oil & Gas Journal,* March 16, 1998.

TABLE 3-1 TYPICAL NGL FEEDSTOCK COMPOSITIONS

	Classification			
	Super-Rich	**Rich**	**Lean**	**Super-Lean**
C1 (mole%)	82.00	87.00	92.00	96.30
C2	9.70	7.90	4.80	2.20
C3	4.50	3.30	1.80	0.80
IC4	0.50	0.46	0.49	0.20
NC4	1.80	1.13	0.87	0.50
IC5	0.50	0.00	0.00	0.00
NC5	0.49	0.21	0.04	0.00
C6+	0.51	0.00	0.00	0.00
BTU/scf	1251.81	1161.77	1105.35	1054.51

If the quantities of these other materials are low, they slightly enhance the heating value of the natural gas and may be left in the gas. If the amount of these materials is such that there may be condensation in the transmission pipeline, or there could be an impact on the suitability of the gas for downstream applications (usually measured in terms of the heating value and/or the related Wobbe number[14]), the pipelines may not purchase the gas from producers without prior treatment to remove the heavy hydrocarbons.

In addition, a large industry has grown to utilize the ethane/propane and heavier hydrocarbons being removed from the natural gas as petrochemical feedstocks. Extraction of NGLs from natural gas can be either profitable or an unavoidable cost of production of natural gas for pipeline transport, depending on the relative price of the hydrocarbons on a heating value and chemical feedstock basis.

Early facilities used circulating heavy hydrocarbon liquids to absorb the lighter components in lean oil absorber plants. In the 1960s, advances in turbomachinery led to newer cryogenic processes being introduced. These turboexpander processes have become the dominant technology.

Technology has evolved to enable the expander plants to have reasonable recoveries and economics when rejecting ethane (leaving it in the gas) while recovering 90% or more of the propane and heavier components. This is necessary to enable such plants to remain in operation when economics favor leaving the ethane in the pipeline gas. There are a number of patented schemes that improve upon the original turboexpander design in order to enable this mode of operation. Many companies offer NGL expander plants including Fluor, Ortloff, IPSI LLC, ProQuip, and ABB Randall.

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[14] The Wobbe number, or Wobbe index, of a fuel gas is found by dividing the high heating value of the gas in BTU per standard cubic foot by the square root of its specific gravity with respect to air. The higher a gases' Wobbe number, the greater the heating value of the quantity of gas that will flow though a hole of a given size in a given amount of time. The usefulness of the Wobbe number is that for any given orifice, all gas mixtures that have the same Wobbe number will deliver the same amount of heat.

3.3 PROCESS DESCRIPTIONS AND DESIGN ASSUMPTIONS

As with the other sections, flow diagrams, stream properties, and an equipment list for both cases is presented below. The multipliers used for installation and indirect costs for the SMR units have also been applied here. After calculation of the equipment cost, a 2.47 factor was used to calculate the Total Installed Cost. For indirect costs, factors developed by NREL to account for engineering costs, construction, contingencies, and fees were applied. Indirect cost factors were then added to the TIC to obtain the Total Project Investment. A summary of the results for both cases can be seen in Table 3-2.

TABLE 3-2 NGL EXPANDER PROJECT COSTS, 2005 BASIS

	Large Scale Plant	Small Scale Plant
Feed Rate (MMSCFD)	400	75
Equipment Cost ($MM)	18.8	4.8
Installed Cost ($MM)	46.4	12.0
Total Plant Investment ($MM)	62.4	16.1
Total Investment/SCFD Feed ($)	0.16	0.21

The differences between the small and large scale cases are not as pronounced as what was seen in the FCC and SMR cases. This is largely due to the relatively small size differences here when compared to the other technologies. NGL expanders of a scale smaller than ~25 MMSCFD are not commercially available; typically, ~75 MMSCFD is considered the low end of what is considered for commercial design. Recovering natural gas liquids from smaller streams would be performed using other technologies, such as straight refrigeration units or Joule-Thomson plants. As can be seen from Table 3-2, there is not a major economy of scale advantage between the high and low end of the commercial plant range.

3.3.1 Small Scale NGL Expander Design

Figure 3-1 shows the PFD for the small scale NGL expander, while Table 3-3 shows the stream conditions. The equipment list for the large scale design can be found in Table 3-4.

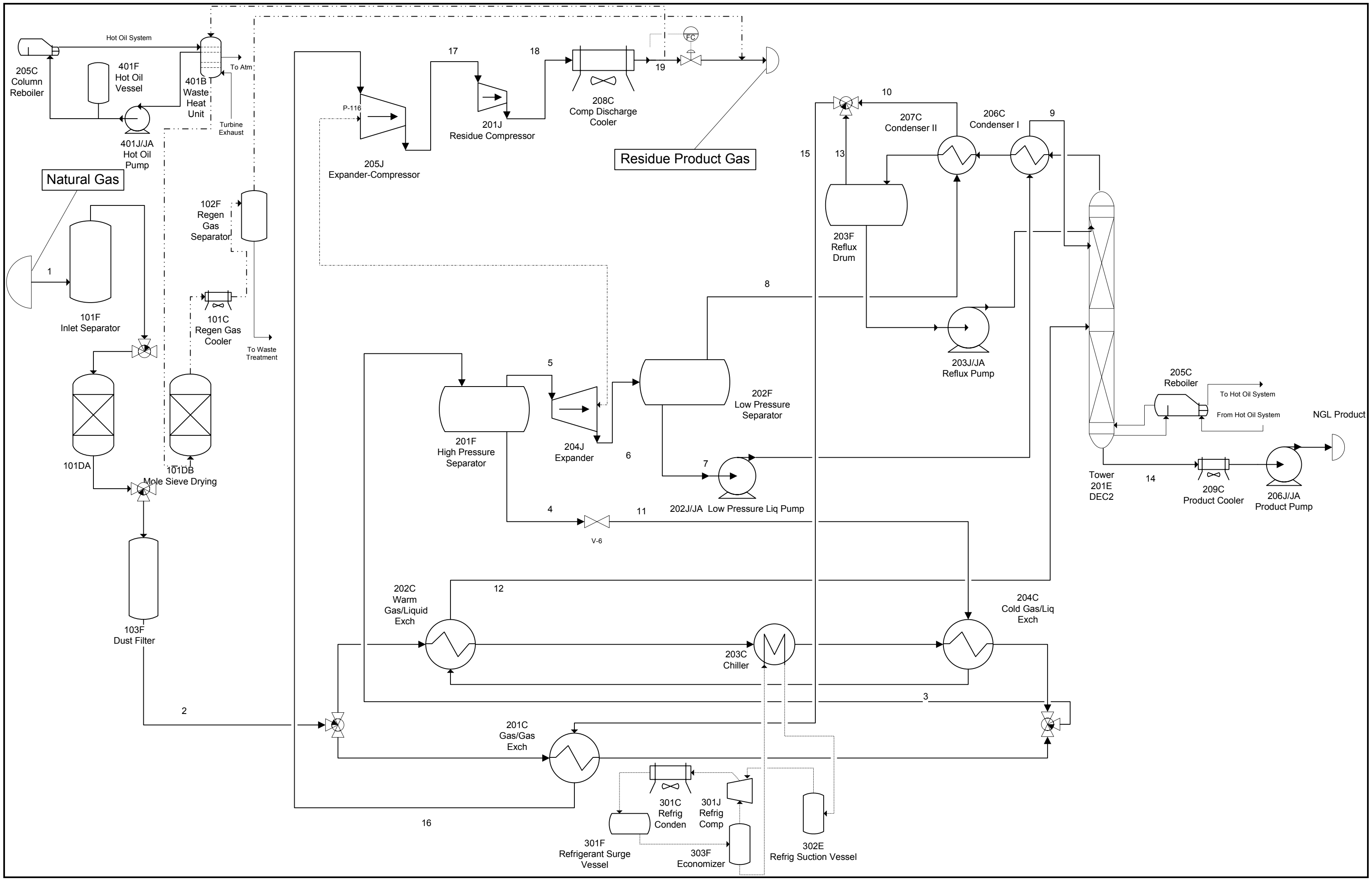

FIGURE 3-1 SMALL-SCALE NGL EXPANDER PROCESS FLOW DIAGRAM

TABLE 3-3 SMALL SCALE EXPANDER STREAM PROPERTIES

Stream No.	1	2	3	4	5	6	7	8	9
Stream ID	Feed Gas	Dry Gas to NGL Unit	To High Pressure Separator	Liq from High Pressure Sep	Vapor from High Pressure Sep	Expander Outlet	Liquid from Low Pressure Sep	Vapor from Low Pressure Sep	Expander Liq to Tower
Methane (lb-mole/h)	6670.1	6670.1	6670.1	756.3	5913.7	5913.7	191.2	5722.6	191.2
Ethane	741.1	741.1	741.1	250.4	490.7	490.7	148.9	341.9	148.9
Propane	436.4	436.4	436.4	235.6	183.9	182.9	134.9	48.0	134.9
i-Butane	82.3	82.3	82.3	60.5	21.9	21.9	19.9	2.0	19.9
n-Butane	148.2	148.2	148.2	116.7	31.6	31.6	29.8	1.7	29.8
i-Pentane	41.2	41.2	41.2	36.2	5.0	5.0	4.9	0.1	4.9
n-Pentane	65.9	65.9	65.9	59.5	6.4	6.4	6.3	0.1	6.3
n-Hexane	35.4	35.4	35.4	33.9	1.5	1.5	1.5	0.0	1.5
n-Heptane	10.7	10.7	10.7	10.5	0.2	0.2	0.2	0.0	0.2
n-Octane	3.3	3.3	3.3	3.3	0.0	0.0	0.0	0.0	0.0
H2O	10.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total lb-mole/h	8,244.7	8,234.6	8,234.6	1,562.9	6,654.9	6,653.9	537.6	6,116.4	537.6
Total lb/h	174,347.5	174,164.8	174,164.8	52,394.3	121,170.5	121,770.5	17,344.3	104,426.2	17,344.3
Std Gas Flow MMscf/d	75.0	74.9	74.9	14.2	60.5	60.5	4.9	55.6	4.9
Total BBL/d	34,273.1	34,260.6	34,260.6	8,156.9	26,103.7	26,103.7	2,815.6	23,288.1	2,815.6
Vapor Fraction	1.0	1.0	0.8	0.0	1.0	0.9	0.0	1.0	0.3
Pressure, psia	1000	980	965	965	965	340	340	340	330
Temperature, F	102.5	102.5	-10.0	-10.0	-10.0	-87.6	-87.6	-87.6	-28.9

Stream No.	10	11	12	13	14	15	16	17	18	19
Stream ID	Expander Vapor from Condenser	Liquid after Valve	High Press Liq to Tower	Vapor from Tower	Tower Bottom	Residue gas to HX	Residue Gas to Exp/Comp	Residue Compr Suction	Residue Compr Discharge	Residue Gas (w/Fuel Gas)
Methane (lb-mole/h)	5722.6	756.3	756.3	947.4	0.1	6670.0	6670.0	6670.0	6670.0	6670.0
Ethane	341.9	250.4	250.4	373.9	25.3	715.8	715.8	715.8	715.8	715.8
Propane	48.0	253.6	253.6	17.4	371.0	65.4	65.4	65.4	65.4	65.4
i-Butane	2.0	60.5	60.5	0.1	80.3	2.0	2.0	2.0	2.0	2.0
n-Butane	1.7	116.7	116.7	0.0	146.5	1.7	1.7	1.7	1.7	1.7
i-Pentane	0.1	36.2	36.2	0.0	41.1	0.1	0.1	0.1	0.1	0.1
n-Pentane	0.1	59.5	59.5	0.0	65.8	0.1	0.1	0.1	0.1	0.1
n-Hexane	0.0	33.9	33.9	0.0	35.4	0.0	0.0	0.0	0.0	0.0
n-Heptane	0.0	10.5	10.5	0.0	10.7	0.0	0.0	0.0	0.0	0.0
n-Octane	0.0	3.3	3.3	0.0	3.3	0.0	0.0	0.0	0.0	0.0
H2O	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total lb-mole/h	6,116.4	1,580.9	1,580.9	1,338.8	779.5	7,455.1	7,455.1	7,455.1	7,455.1	7,455.1
Total lb/h	104,426.2	52,394.3	52,394.3	27,219.8	42,518.7	131,646.0	131,646.0	131,646.0	131,646.0	131,646.0
Std Gas Flow MMscf/d	55.6	14.4	14.4	12.2	7.1	67.8	67.8	67.8	67.8	67.8
Total BBL/d	23,288.1	8,156.9	8,156.9	5,745.6	5,226.9	29,033.7	29,033.7	29,033.7	29,033.7	29,033.7
Vapor Fraction	1.0	0.4	0.7	1.0	0.0	1.0	1.0	1.0	1.0	1.0
Pressure, psia	335	370	365	340	350	335	325	435	1010	1000
Temperature, F	-36.3	-46.0	90.0	-43.0	199.1	-38.1	90.0	137.9	287.2	120.0

TABLE 3-4 SMALL-SCALE NGL EXPANDER EQUIPMENT LIST

						Design		Operating							
Item No	**Description**	**Type**	**Quantity Per Train**	**Size**	**Weight**	**P**	**T**	**P**	**T**	**Power Useage**	**Materials**	**Price, total (unistalled)**	**Total Installed Cost**	**Quote Source**	**Comments**
					lbs	PSIG	ºF	PSIG	ºF	(No.) HP		2002 Cost Index $US	(US $)		
Reactors															
101D	Mole Sieve Dryer	Packed Bed	2	66" D x 26' T/T	52,000	1,100	600	1,000	100		CS	$51,375		Trimenic Design	requires 19500 lb of mol sieve
Total												**$51,375**			
Exchangers															
201C	Gas/Gas Exchanger	Shell/Tube	1	UA = 703000		1,100	-20	1,000	-10		Aluminum	$95,000		KBR Design	9.4 MMBTU/hr
202C	Warm Gas/Liquid Exchanger	Shell/Tube	1	UA = 132000		1,100	-20	1,000	-10		Aluminum	$114,000		KBR Design	2.4 MMBTU/hr
203C	Chiller	Shell/Tube	1	UA = 106000		1,100	-20	1,000	-10		Aluminum	$85,500		KBR Design	1.6 MMBTU/hr
204C	Cold Gas/Liquid Exchanger	Shell/Tube	1	UA = 189000		1,100	-20	1,000	-10		Aluminum	$120,000		KBR Design	4.2 MMBTU/hr
205C	Reboiler	Shell/Tube	1	200 sq. ft.		470	-50	350	200		CS	included in 201E		KBR Design	5.8 MMBTU/hr
206C	Condenser I	Shell/Tube	1	UA = 21000		470	-100	350	50		Aluminum	$78,000		KBR Design	1 MMBTU/hr
207C	Condenser II	Shell/Tube	1	UA = 94000		470	-100	340	-40		Aluminum	$76,000		KBR Design	3.2 MMBTU/hr
208C	Comp. Discharge Cooler	Air Fin	1	4290 sq ft		1,100	300	1,000	290	43	CS	$100,000		KBR Design	14 MMBTU/hr
209C	Product Cooler	Air Fin	1	1270 sq ft		470	250	350	200	12	CS	$30,000		KBR Design	2.8 MMBTU/hr
Total												**$698,500**			
Columns, Vessels & Tanks															
101F	Inlet Filter Separator	Vertical	1			1100	150	1000	100		CS	$27,000		KBR Design	
103F	Dust Filter	Vertical	1			1100	200	1000	100		CS	$20,000		KBR Design	
201F	High Pressure Separator	Horizontal	1	48" x 10' T/T	12,000	1100	-20	1000	-10		CS	$148,540		KBR Design	Includes mesh pad
202F	Low Pressure Separator	Horizontal	1	42" x 8' T/T	3,000	470	-100	350	-85		SS	$98,522		KBR Design	Includes mesh pad
203F	Reflux Drum	Horizontal	1	48" x 10' T/T	4,700	470	-50	340	-40		CS	included in 201E		KBR Design	Includes mesh pad
201E	Deethanizer	Vertical	1	52" x 40' T/T	26,000	470	-50	350	50		LTCS	$682,073		KBR Design	Packed with 1.5" Pall Rings
Total												**$976,135**			
Pumps, Compressors & Blowers															
204J	Expander	Turbine	1	1217 HP		400	-100	350	-85	1,217	CS/SS	$565,685		KBR Design	
202J	Low Pressure Liquid Pump	Centrifugal	2	70 gpm x 50 psig		470	-100	350	-85	3	SS	$108,902		KBR Design	
203J	Reflux Pump	Centrifugal	2	95 gpm x 50 psig		470	-50	340	-40	5	CS	$48,099		KBR Design	
205J	Expander Compressor	Turbo-Expander	1	1217 HP		1,100	150	435	130	1,217	CS	included with 204J		KBR Design	
201J	Residue Compressor	Centrifugal	1	4113 HP		1,100	150	1,000	290	4,113	CS	$2,375,879		KBR Design	
206J	Product Pump	Centrifugal	2	98 HP		470	250	350	200	98	CS	$40,000		KBR Design	
Total												**$3,138,565**			
Other															
	Refrigeration System									180		Included in 203C		KBR Design	
	Hot Oil System									21		Included in 201E		KBR Design	
SUBTOTAL												**$4,864,575**			
TOTAL PROJECT COST													**$12,015,501**		

The feed gas, stream 1, at 1000 psia enters the inlet filter separator (101F) where any entrained liquids and solids are removed. The gas, free of entrained liquids, flows to the mole sieve dryer (101DA or 101DB) where the water is adsorbed on the mole sieves, reducing the water content of the gas to less than 1 ppm. The mole sieve beds are regenerated by bringing turbine exhaust gas (which is predominantly methane) from the residue compressor (201J) through the waste heat unit (401B) where a portion of the gas from the NGL recovery section is heated to 550°F prior to flowing through the switching valves into the mole sieve dryer. The gas then enters the dust filter (103F) where any mole sieve fines are removed. From the dust filter, the dry gas, stream 2, enters the NGL recovery section of the plant.

The dehydrated gas is split to act as a heat exchange medium for a number of different streams. 46% of the gas goes through the gas/gas exchanger (201C), where it is cooled to -25°F against the residue gas. The remainder of the gas flows through the warm gas/liquid exchanger (202C), the chiller (203C), and the cold gas/liquid exchanger (204C) to cool the stream to 4°F. These two split streams are recombined, and the combined stream, stream 3, enters the high pressure separator (201F) where the cold liquids are separated from the gas. The cold liquids, stream 4, leave the separator and are reduced in pressure across a valve to reduce the temperature to -45°F.

Gas stream 5, from the high pressure separator (201F) is expanded almost isentropically in the expander (204J) to 340 psia. This gas stream, stream 6, enters the low pressure separator (202F) where the liquids are separated from the gas. These cold liquids, stream 7, are pumped by the low pressure liquid pump (202J, JA) through condenser I (206C) to the deethanizer (201E). The gas from the low pressure separator, stream 8, flows through condenser II (207C) before being recombined with the deethanizer overhead, stream 10.

The deethanizer (201E) is a packed tower which separates the propane and heavier components from the methane and ethane. Heat is supplied to the deethanizer by a hot-oil-heated reboiler (205C). The overhead vapors from the deethanizer are partially condensed in condenser I (206C) and condenser II (207C) using the cold gases and liquids from the low pressure separator (202F). The liquids are separated in the reflux drum (203F) and pumped back to the deethanizer by the reflux pump (203J,JA).

Stream 13 is combined with stream 10 from the low pressure separator (202F) and used for cooling in the gas/gas exchanger (201C). This gas, stream 16, is partially re-compressed by the expander compressor (205J), which uses the power generated by the expander (204J). The gas, stream 17, is further compressed to pipeline pressure by the gas turbine driven residue compressor (201 J). The residue compressor discharge, stream 18, is cooled by the compressor discharge cooler (208C) prior to entering the pipeline.

External refrigeration is provided to the chiller (203C) by a propane refrigeration system. The gas engine driven refrigerant compressor (301J) compresses the propane to 245 psia after which it is condensed using air in the refrigerant condenser (301C). The liquid propane then goes to the refrigerant surge vessel (301F) and on to the economizer (303F) where it is flashed. The economizer serves to reduce the horsepower requirements of the refrigeration system. The vapors from the economizer are returned interstage into the refrigerant compressor. The liquid from the economizer goes to the chiller (203C) where it vaporizes to provide cooling for the inlet gas to the NGL recovery section of the expander plant. The vapors from the chiller go through the refrigerant suction vessel (302F) to remove any entrained liquid and are then returned to the suction of the refrigerant compressor.

A hot oil system provides energy for the regeneration of the dryers. Heat for the mole sieve regeneration gas and the hot oil is provided by the turbine exhaust from the residue gas compressor (201J). This heat is recovered in the waste heat unit (401B). Hot oil is pumped through this unit by the hot oil pump (401J, JA) to the deethanizer column reboiler (205C). Hot oil is returned from the reboiler to the suction of the hot oil pump.

Design Assumptions

The small scale NGL design was based on a study performed by M. W. Kellogg for GRI[15]. A self-contained, turbo-expander plant for the recovery of NGL from natural gas was designed and costed. Standard sizing techniques were used to estimate the size and weight of process equipment. Telephone estimates were received from vendors for most equipment. Bulks for the process units were determined using a Kellogg-proprietary estimating program with adjustments for the use of skid-mounted units. The costs were updated to 2005 using the Chemical Engineering Plant Cost Index.

The design presented here assumes a stand-alone facility that could be used for most NGL expander designs. The feed natural gas entering the plant is assumed to require additional treating, which is why both drying and particulate removal units have been included. While these steps may not always be necessary for a unit placed on a common natural gas header, including them gives the plant extra flexibility to handle a variety of feeds.

Heating and cooling loads in an expander facility may vary considerably based on the overall feed rate, pipeline quality requirements, and the level of fractionation desired. For this design, both hot oil and refrigeration systems have been included to provide additional plant flexibility. These units are typically seen more often in small units rather than large ones which have more stable plant operations.

3.3.2 Large Scale NGL Expander Design

Figure 3-2 shows the PFD for the small scale NGL expander, while Table 3-5 shows the stream conditions. The equipment list for the large scale design can be found in Table 3-6.

The large scale process for NGL recovery is very similar to that of the smaller scale systems using the same turbo-expander technology. The major difference is heat integration—the larger scale systems are able to afford additional heat exchanger systems to decrease the thermodynamic penalties related to cooling the stream. This adds to the overall complexity of the design; while increasing capital costs, the total investment per SCFD of feed decreases.

Inlet gas enters the plant at 120°F and 900 psia (stream 1). If the inlet gas contains a concentration of sulfur compounds which would prevent the product streams from meeting specifications, the sulfur compounds are removed by appropriate pretreatment of the feed gas. In addition, the feed stream is usually dehydrated upstream to prevent hydrate (ice) formation under

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[15] Evaluation of Natural Gas Processing Technologies, Topical Report, Task 3, Volume II, M. K. Changela, et al, The M. W. Kellogg Company,(1990).

cryogenic conditions. Solid desiccants are typically used for this purpose and are not shown here.

The feed is cooled in several heat exchangers with cool residue gas at –86°F, bottom liquid product at 62°F, demethanizer reboiler liquids at 39°F, and demethanizer side reboiler liquids at -32°F. The cooled stream 3 enters high pressure separator (V-100) at -22°F and 885 psia where the vapor (stream 5) is separated from the condensed liquid (stream 4).

The vapor from separator V-100 (stream 5) enters the turbo-expander (K-100) in which mechanical energy is extracted from this portion of the high pressure feed. This unit expands the vapor isentropically from a pressure of 885 psia to a pressure of 312 psia, with the work expansion cooling the expanded stream 6 to a temperature of approximately -112°F. Typical commercially available expanders are capable of recovering on the order of 80-85% of the work theoretically available in an ideal isentropic expansion. The work recovered is used to drive the centrifugal expander compressor (K-101). The expanded and partially condensed stream 6 is further flashed in the low-pressure separator (V-101).

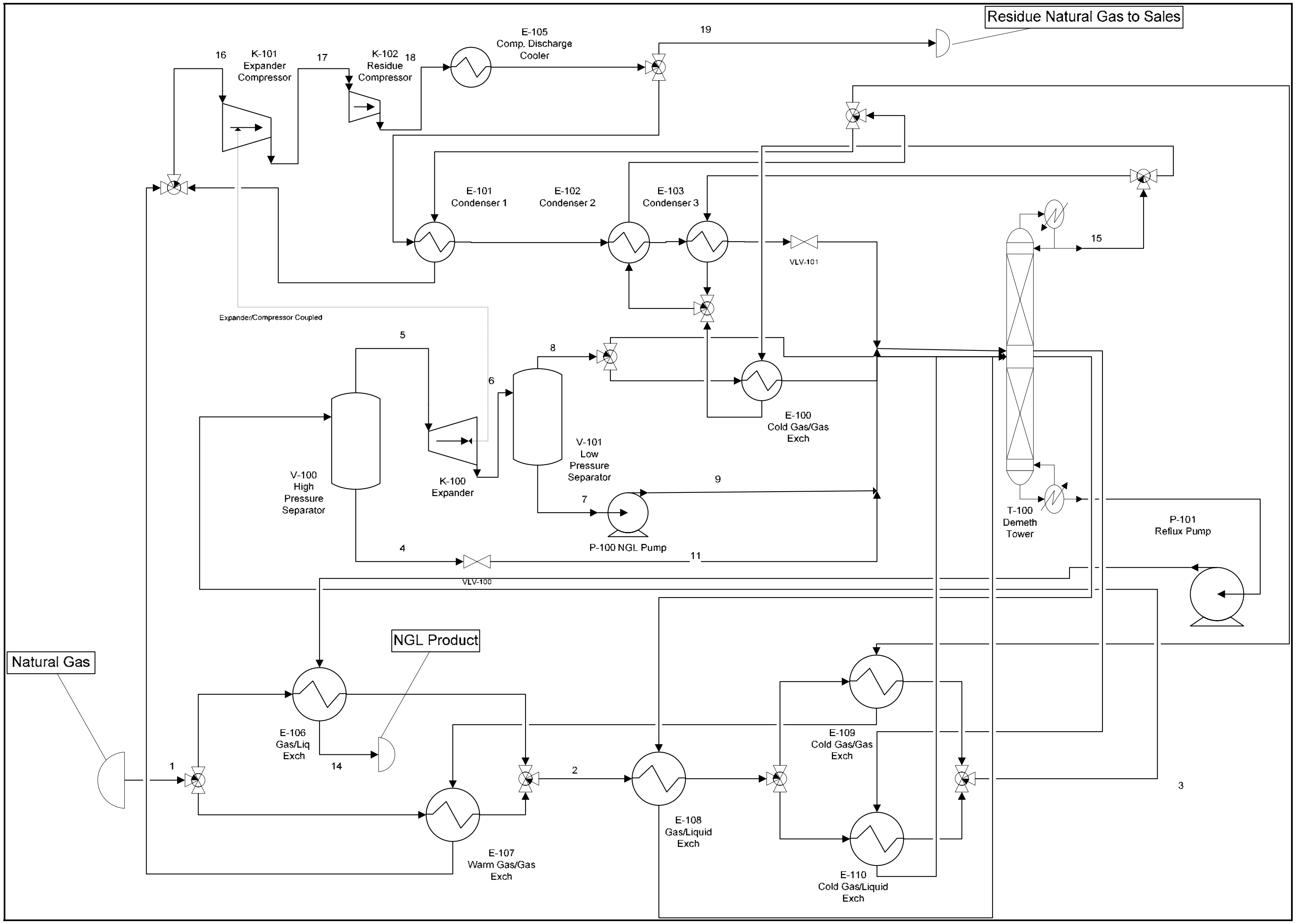

FIGURE 3-2 LARGE-SCALE NGL EXPANDER PROCESS FLOW DIAGRAM

TABLE 3-5 LARGE SCALE EXPANDER STREAM PROPERTIES

Name	1	2	3	4	5
Methane [lbmole/h]	36014.15	36014.15	36014.15	6272.595	29755.73
Ethane	4260.211	4260.211	4260.211	2042.24	2223.505
Propane	1976.387	1976.387	1976.387	1436.301	523.1776
n-Butane	1668.949	1668.949	1668.949	1481.185	196.1916
Water	0	0	0	0	0
Molar Flow [lbmole/h]	43919.7	43919.7	43919.7	11221.1	32698.6
Vapor Fraction	1.00	1.00	0.82	0.00	1.00
Temperature [F]	120.0	72.0	-22.0	-38.6	-38.6
Pressure [psia]	900	900	900	885	885
Mass Flow [lb/h]	890,033.3	890,033.3	890,033.3	311,080.5	578,952.9
Std Ideal Liquid Flow [barrels/d]	179,967.1	179,967.1	179,967.1	53,441.8	126,525.3
Std Gas Flow [MMSCFD]	400.0	400.0	400.0	102.2	297.8

Name	6	7	8	9	11
Methane [lbmole/h]	29755.73	1262.272	28487.46	1262.272	6272.595
Ethane	2223.505	958.1792	1252.852	958.1792	2042.24
Propane	523.1776	461.8768	89.4894	461.8768	1436.301
n-Butane	196.1916	189.3408	0	189.3408	1481.185
Water	0	0	0	0	0
Molar Flow [lbmole/h]	32698.6	2868.8	29829.8	2868.8	11221.1
Vapor Fraction	0.91	0.00	1.00	0.00	0.40
Temperature [F]	-112.0	-112.0	-112.0	-111.9	-83.5
Pressure [psia]	312	312	312	324	325
Mass Flow [lb/h]	578,952.9	80,332.8	498,620.1	80,332.8	311,080.5
Std Ideal Liquid Flow [barrels/d]	126,525.3	14,207.6	112,317.6	14,207.6	53,441.8
Std Gas Flow [MMSCFD]	297.8	26.1	271.7	26.1	102.2

Name	14	15	16	17	18	19
Methane [lbmole/h]	205.3818	45039.73	1345.113	1352.006	45045	35810.74
Ethane	4044.442	271.8696	3301.641	3301.884	271.9014	216.1614
Propane	1974.825	0	1202.45	1202.538	0	0
n-Butane	1674.652	0	937.503	937.572	0	0
Water	0	0	0	0	0	0
Molar Flow [lbmole/h]	7899.3	45311.6	6793.5	6794.0	45316.9	36026.9
Vapor Fraction	0.00	1.00	0.00	0.46	1.00	1.00
Temperature [F]	115.0	-155.1	-47.5	38.0	288.5	90.0
Pressure [psia]	600	300	314	314	900	900
Mass Flow [lb/h]	309,317.9	730,573.2	228,587.0	228,570.1	730,655.8	580,871.3
Std Ideal Liquid Flow [barrels/d]	47,364.8	166,802.7	37,646.1	37,644.6	166,821.9	132,623.4
Std Gas Flow [MMSCFD]	71.9	412.7	61.9	61.9	412.7	328.1

TABLE 3-6 LARGE-SCALE NGL EXPANDER EQUIPMENT LIST

						Design		Operating							
Item No	**Description**	**Type**	**Quantity Per Train**	**Size**	**Weight**	**P**	**T**	**P**	**T**	**Power Useage**	**Materials**	**Price, total (unistalled)**	**Total Installed Cost**	**Quote Source**	**Comments**
					lbs	PSIG	ºF	PSIG	ºF	(No.) HP		2002 Cost Index $US	(US $)		
Exchangers															
E-100	Cold Gas/Gas Exchanger	Shell/Tube	1	Area = 10,120 sq. ft.		350	-155	312	-112		Aluminum	$179,197		Chart Industries	9.2 MM Btu/h
E-101	Condenser I	Shell/Tube	1	Area = 16,145 sq. ft		1,000	-70	300	-60		Aluminum	$174,602		Chart Industries	17.2 MM Btu/h
E-102	Condenser II	Shell/Tube	1	Area = 15,725 sq. ft		1,000	-115	300	-110		Aluminum	$222,847		Chart Industries	14.8 MM Btu/h
E-103	Condenser III	Shell/Tube	1	Area = 17,990 sq. ft.		1,000	-150	300	-150		Aluminum	$192,981		Chart Industries	8.0 MM Btu/h
E-105	Compressor Discharge Cooler	Shell/Tube	1	Area = 17,480 sq. ft.		1,000	300	900	288		CS	$229,740		Chart Industries	92.0 MM Btu/h
E-106	Gas/Liquid Exchanger	Shell/Tube	1	Area = 24,025 sq. ft.		1,000	150	900	120		CS	$261,903		Chart Industries	12.4 MM Btu/h
E-107	Warm Gas/Gas Exchanger	Shell/Tube	1	Area = 8,220 sq. ft.		1,000	150	900	120		CS	$218,253		Chart Industries	15.2 MM Btu/h
E-108	Gas/Liquid Exchanger	Shell/Tube	1	Area = 22,245 sq. ft.		1,000	100	900	72		CS	$275,688		Chart Industries	16.4 MM Btu/h
E-109	Cold Gas/Gas Exchanger	Shell/Tube	1	40 MMBTU/hr		1,000	-100	900	-20		Aluminum	$252,714		Chart Industries	40.0 MM Btu/h
E-110	Cold Gas/Liquid Exchanger	Shell/Tube	1	Area = 13,324 sq. ft.		1,000	-50	350	-20		Aluminum	$275,688		Chart Industries	23.6 MM Btu/h
Total												**$2,283,613**			
Columns, Vessels & Tanks															
T-100	Demethanizer	Tray Column	1	78" x 85' T/T		400	-160/60	300	-155/30		LTCS	$1,033,829		Dickerson Trayer	24 inch tray spacing
V-100	High Pressure Separator	Horizontal	1	72" x 15' T/T	68,000	1000	-40	900	-22		CS	$225,145		Dickerson Trayer	Liquid Volume = 10,000 gal
V-101	Low Pressure Separator	Horizontal	1	63" x 12' T/T	102,000	400	-120	312	-112		SS	$149,331		Dickerson Trayer	Liquid Volume = 2,735 gal
Total												**$1,408,305**			
Pumps, Compressors & Blowers															
P-100	NGL Pump	Centrifugal	2	400 HP		650	-120	350	-100	400	SS	$275,688		Union	1448 gpm
P-101	Reflux Pump	Centrifugal	2	12 HP		350	-120	350	115	12	SS	$121,762		Union	360 gpm
K-100	Expander	Turbine	1	5180 HP		1,000	-120	900	-85	5,180	CS	$2,828,427		Atlas Copco	
K-101	Expander Compressor	Turbo-Expander	1	5180 HP		1,000	150	314	-50	5,180	CS	included with K-100		Atlas Copco	
K-102	Residue Compressor	Centrifugal	1	26,624 HP		1,000	150	900	288	26,624	CS	$11,879,394		Hanover	
Total												**$15,105,271**			
SUBTOTAL												**$18,797,189**			
TOTAL PROJECT COST													**$46,429,057**		

The vapor (stream 8) from V-101 is divided into two streams. About 60% of the total vapor, passes through the cold gas/gas exchanger (E-100) against a portion of the cold distillation stream, resulting in further cooling and partial condensation. The further cooled and partially condensed stream is then supplied to the distillation column at an upper mid-column feed point.

The remaining 40% of the vapor from V-101 is fed to the demethanation tower (T-100) at a second upper mid-column feed point. The condensed liquid (stream 9) from V-101 is pumped in the NGL pump (P-100) prior to feeding the distillation column at a third upper mid-column feed point.

A portion of the high-pressure residue gas is withdrawn from the main residue flow (stream 18) to become the top distillation column feed. The recycle gas stream passes through a series of heat exchangers with a portion of the cool residue gas where it is cooled to -148°F. As the stream is expanded to the demethanizer's operating pressure of 324 psia, it is cooled to a temperature of approximately -158°F.

The cooling and partial condensation of a portion of the vapor phase leaving the work expansion machine results in additional liquids being supplied to the fractionation tower at the upper mid-column feed position, whereupon these liquids act as reflux on the vapor rising up the tower. This in turn reduces the amount of vapor to be rectified by the top reflux stream (and also creates a leaner vapor composition), allowing a corresponding reduction in the amount of top reflux feed required to achieve the desired C2+ component recovery level.

Design Assumptions

The large scale NGL design was based on a study performed by GTI for DOE. A turbo-expander plant for the recovery of NGL from natural gas was designed and costed. A HYSYS simulation model was used to size equipment. Costs were verified by Trimeric Corporation using vendor quotes and PDQ$ costing program.

Large scale units have stable feed rates and qualities, since they are typically built in an area with a very steady feed supply. In addition, product demand is usually locked in and integrated with petrochemical facilities. The design flexibility incorporated into the small scale unit, processes for feed treating and additional heating/cooling, is typically not necessary in a large scale unit.

3.4 OPERATING COSTS AND UTILITY REQUIREMENTS

The basis for the operating costs and utility requirements are shown in Table 3-7.

TABLE 3-7 NGL EXPANDER OPERATING COSTS AND UTILITY REQUIREMENTS

Operating Variable	Large Scale	Small Scale	Notes
Cooling Water (gal/hr)	52,568	10,000	Interstage compressor requirements
Catalyst *Mol Sieve (lb)*	N/A	19,500	~$2.50/lb, lifespan will vary
Electricity Use (MW)	27.9	5.1	Product compression is main cost

Besides the items listed above, natural gas is also used for various heating requirements, such as reboilers and hot oil system reheating. Typically, the fuel gas produced from the expander system is used for this service, so no external natural gas is purchased specifically for plant use.

As mentioned earlier, the feed natural gas may require additional treating, which is why both drying and particulate clean-up units have been included in the small scale design. Large scale units which have more stable and reliable feed sources may have treating upstream of the facility and may not require these units to be present. Information provided by vendors for the large scale designs did not include these minor pieces of equipment, which is why no molecular sieve requirement is listed for the large scale unit.

When attempting to draw conclusions based on the scaling factors for plant sizes and costs addressed here, it should be recognized that the plant ranges considered are very different. This is important not only because of the sheer magnitude difference, but also for where the plants are situated in the range of commercial sizes. For example, comparing plants at the very low and high end of their respective commercial ranges will differ from a comparison between plants at other points along the commercial spectrum. This is compounded by the fact that the "commercial range" is a moving target based on technology advances and the industry where the technology is used.

Most technologies have some asymptotic economy of scale when capital investment is compared to plant size. This is demonstrated in Figure 4-1 below, an evaluation of the economy of scale advantage for large scale SMR technology.

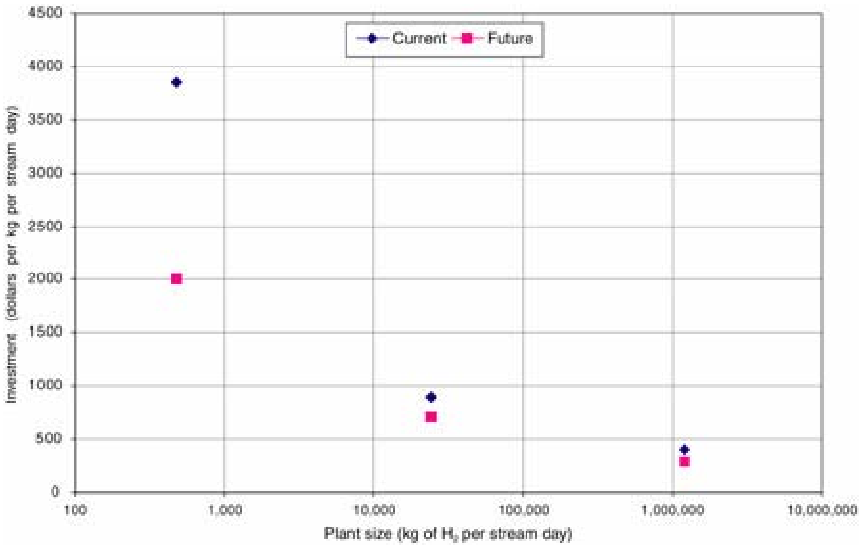

FIGURE 4-1 SMR PROJECT COSTS, BROAD SIZE RANGE[16]

As can be seen from Figure 4-1, the economy of scale diminishes as the plant size approaches the high end of the commercial scale. When comparing two different plant sizes, it is important to consider at what points along this curve the plants are situated. While a power relationship is helpful in performing sizing analysis regardless of where the two plants are in relation to one another, more accuracy can be gained if the power is modified based on the relative position to each other on the curve.

For the technologies evaluated in this study, different commercial size ranges were evaluated. The ranges are:

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[16] The Hydrogen Economy: Opportunities, Costs, Barriers, and R&D Needs (2004), National Academy of Engineering, Board on Energy and Environmental Systems

- FCC: Very low end commercial to mid-to-high end commercial. No commercial scale FCC units are built today below ~10,000 BPSD, due to the poor economics of units this size.
- SMR: Very low end niche commercial unit to high end commercial unit for industrial purposes. The small scale unit would be used for hydrogen production for a small fuel cell application, while the large scale unit would be appropriate for a large industrial application.
- NG Expander: Low-to-mid commercial to high end commercial. The range of commercial applications for NG expanders is lower than other technologies, due to technology breakpoints at the low end and a switch to multiple trains on the high end. Plant scale evaluations were limited to single train systems.

Although the size ranges that were considered are different, an attempt was made to determine if a simple power relationship could be used to predict the cost of each small scale plant. The following basis was used:

$$\text{Small Scale Cost} = \text{Large Scale Cost} * \left(\frac{\text{Small Scale Capacity}}{\text{Large Scale Capacity}} \right)^{0.7}$$

The results of this analysis showed that at a minimum, an order of magnitude predictive basis is possible using this relationship for all technologies evaluated, regardless of the differences in size. In most cases, using a power basis provided a cost estimate within a level of acceptability (+/- 30%) for this level of detail. The analysis performed is detailed in Table 4-1 below.

TABLE 4-1 DESIGN RATIOS AND PREDICTED COSTS, ALL TECHNOLOGIES

	FCC	SMR	NG Expander
Large/Small Scale Size Ratio (Feedstock)	25	7,000	5.3
Large/Small Scale Cost Ratio	9.4	1,000	3.8
Size Ratio/Cost Ratio	2.7	7	1.4
Predicted Small Cost, Power Basis ($MM)	20.2	0.75	19.3
Actual Small Cost, Design Basis ($MM)	20.5	0.39	16.1
Percent Difference	1.5%	92.3%	19.8%

Although applying a power relationship to SMR reforming technology had the worst predictive capacity, the resulting estimate is still very close on an absolute basis. It was expected that this technology would have the most difficulty in being able to predict by a power relationship because of the considerable difference in size relative to the other technologies chosen.

This power relationship can be used for many different types of technologies to provide an acceptable cost estimate for planning purposes. However, two major items should be kept in mind as this approximation is applied. First, the level of technological maturity will have an impact on how the costs scale when the plant size is changed. All operations being analyzed in this study are commercially mature, with little future cost reduction expected from technological breakthroughs. The impacts of future cost reductions due to technical improvements must be considered separately outside of this analysis. Secondly, the power that should be used when comparing two units will vary based upon their relative size. Comparing large scale units in the "flat" portion of Table 4-1 will be better accomplished with a larger exponent (reflecting a more linear relationship), while a smaller exponent should be used when comparing units that have a more asymptotic relationship.

A few general conclusions can be made regarding the cost relationship as technologies are scaled:

- The power relationship that should be used depends largely on the size range that is being considered. For example, a very large difference in sizes (SMR technology) will usually be much more difficult to accurately predict than smaller differences.
- Economies of scale are usually greater when technologies contain a novel or expensive main piece of equipment, such as the catalytic furnace in an SMR unit or a reactor/regenerator unit in an overall FCC design. Established technologies, such as refrigeration and fractionation in the NG Expander designs, are more easily predicted as to their impact when sizes change.
- Many technologies have a breakpoint where the type of design changes due to size factors. This is true for all technologies considered in this study, as well as for many other technologies such as oxygen production units. As designs get close to this technology breakpoint, the capital cost per unit of feed changes considerably.

The power relationships explored here could be applied to technologies of particular interest to NREL, such as biomass gasifiers and tar crackers. Since neither of these technologies is mature, estimates for large scale applications will need to take into account the cost reductions expected as the technology progresses toward commercialization. Scaling from the small scale units currently operated or being explored by NREL to large, commercial units would likely require a smaller power relationship than the one used in Table 4-1. Any attempt to estimate unit costs from scaling factors should be checked with a preliminary (+/- 30%) design and cost estimate, using design software and vendor quotes. This step is important to provide a reality check for units that do not have widespread commercial application, and to provide additional data to support theoretical predictions.

After discussions with NREL, it was decided that Subtask 1.1, Technology Evaluation, would focus on Chemical Process Technologies. The final list of technologies chosen was:

- Refinery Processes
- Claus Sulfur Recovery
- Natural Gas Reforming to Hydrogen
- Methanol from Natural Gas
- Ammonia Production
- Gas-To-Liquids
- Natural Gas Liquids Expanders
- Cryogenic Oxygen Plants
- Ethylene and Olefins Plants

Information about all of these technologies, including history, technology types, a description of the main technologies used, and commercial process ranges, was produced as the deliverable for Subtask 1.1. The main content of that deliverable is presented in this Appendix.

After a review of Subtask 1.1, three technologies, Fluidized Catalytic Cracking of refinery gas oil, Natural Gas Reforming, and Natural Gas Liquids Expanders, were chosen for further investigation. These technologies were selected due to how the relationship between cost and equipment size at both a large and small scale is useful for application toward other technologies of interest to NREL.

A.1 REFINERY PROCESSES

A.1.1 Introduction

As with many commodity chemical processes, petroleum refineries have increased in scale considerably since the first ones were built in the early 20th century. Economies of scale have played a large part in defining the current make-up of petroleum refineries in the US today. Due in part to the expansion of distribution facilities, environmental regulations, and the removal of price controls in 1981, many of the very small scale (less than 10,000 barrels per day (BPD)) refineries have shutdown since 1980. The total number of refineries has decreased from over 300 in 1980 to 144 in 2004[17,18].

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[17] 2004 data from the US Department of Energy EIA-820 Refinery Capacity Survey, January 1, 2004.

[18] Historic information from the DOE commissioned report by the National Petroleum Council, "US Petroleum Refining, Assuring the Adequacy and Affordability of Cleaner Fuels", June 2000.

A.1.2 Overall Refinery Sizes

While considerable range still exists in US petroleum refineries, many very small scale refineries are filling either a geographic or specialty product niche. Of the 144 refineries, only 11 (8%) have a capacity under 10,000 BPD, while 62 (43%) have capacities over 100,000 BPD. Figure A-1 below has a breakdown of current refinery capacities:

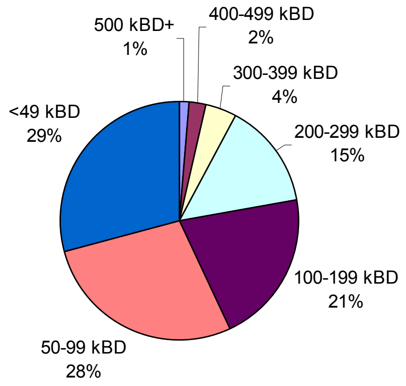

FIGURE A-1 2004 US REFINERY SIZES

While refineries less than 1,000 BPD have historically been in commercial operation, no refineries of this size are currently in operation in the US. The difference between the largest and smallest 2004 nameplate capacities is a factor of nearly 300. Table A-1 below shows the largest and smallest US refinery capacities as of January 1, 2004[19].

Table A-1 Largest/Smallest US Refineries
(Nameplate Calendar Day Capacity, January 2004)

Company	Location	Capacity (BPD)	Major Outputs
LARGEST			
ExxonMobil	Baytown, TX	557,000	Fuels, Chemicals, Lubricants
Hovensa LLC	Kings Hill, USVI	495,000	Fuels, Chemicals, Lubricants
ExxonMobil	Baton Rouge, LA	493,500	Fuels, Chemicals, Lubricants
SMALLEST			
Foreland Refining	Eagle Springs, NV	1,925	Asphalt, Bunker Fuels
Tenby Inc.	Oxnard, CA	2,800	Asphalt, Bunker Fuels
Silver Eagle Refining	Evanston, WY	3,000	Asphalt, On-Road Fuels

With every refinery having a different process flow and complexity level based on the crude types fed and the product slate chosen, it is difficult to make general comparisons between all refineries. A 500,000 BPD refinery making a variety of products will look very different from a 2,000 BPD refinery that produces asphalt and bunker fuels. Large refineries will tend to have a greater level of complexity to optimize process economics by upgrading low quality crudes into

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[19] All refinery process data from the US Department of Energy EIA-820 Refinery Capacity Survey, January 1, 2004.

valuable clean fuels. Smaller refineries cannot afford this level of complexity, focusing on a very limited input and output slate.

A.1.3 Process Unit Sizes

In order to provide comparisons on a more equal footing, petroleum refineries have been broken down by process unit type. Each of these process units are complicated processes by themselves, with many having been scaled considerably. Process size ranges are provided for each unit type. Please note that the capacities are for individual units at the plant and not the total amount of capacity available at the site. For example, although ExxonMobil's Baytown refinery has 557,000 BPD of atmospheric distillation capacity, this is performed over a number of individual units.

Atmospheric and Vacuum Crude Distillation

The first step of any petroleum refinery is to feed crude oil into a distillation column to obtain the rough product cuts that will be further refined and blended downstream. Most initial distillation is done at atmospheric conditions. When feeding a heavy crude slate, bottoms from atmospheric distillation units are sometimes sent to a vacuum crude tower for further component separation.

Table A-2 Atmospheric/Vacuum Crude Distillation Unit Capacities

Company	Location	Capacity (BPD)
LARGE		
ExxonMobil	Baytown, TX	90,000 to 120,000 (atm.)
ChevronTexaco	Pascagoula, MS	70,000 to 100,000 (vac.)
SMALL		
Foreland Refining	Eagle Springs, NV	1,925 (atm)
Wyoming Refining	Newcastle, WY	1,800 (atm)

Alkylation

This process consists of the reaction of isobutane with a mixed light olefinic (usually C3 and C4) stream to produce a high octane gasoline blending component. The resulting product is usually blended to make premium, 90 to 93 octane, gasoline. This reaction occurs at cold temperatures and low pressures, using stirred sulfuric acid as a reaction catalyst.

Table A-3 Alkylation Unit Capacities

Company	Location	Capacity (BPD)
LARGE		
BP	Texas City, TX	20,000 to 30,000
SMALL		
Montana Refining	Great Falls, MT	800

Aromatics

In general, aromatics units tend to be pair with more complex refineries that have both reforming capacity and a strong market for aromatics products (benzene, toluene, and xylenes). Large refineries that are paired with olefins plants also usually possess some sort of aromatics processing capacity. Raw feed from refinery reformers or heavy sections of olefins plants are sent to aromatics processing units for extraction. This is usually a physical conversion, which consists of solvents, zeolite adsorption, and distillation.

Table A-4 Aromatics Unit Capacities

Company	Location	Capacity (BPD)
LARGE		
BP	Texas City, TX	45,000
SMALL		
Alon USA LP	Big Spring, TX	1,000

Fluidized Catalytic Cracking

A standard process in many refineries is the upgrading of gas oil to gasoline. FCC units have been present in US refineries for over 50 years, and are considered a very mature technology. In this process, a fluidized catalyst reacts with an inlet gas oil stream at high pressure to produce a predominantly unsaturated product stream suitable for gasoline blending. The catalyst is separated from exit gases in a cyclone, regenerated in a separate reactor, and then reintroduced into the process reactor.

Table A-5 Fluidized Catalytic Cracking Unit Capacities (Fresh Feed)

Company	Location	Capacity (BPD)
LARGE		
ExxonMobil	Baytown, TX and Baton Rouge, LA	70,000 to 110,000
SMALL		
Montana Refining	Great Falls, MT	2,400

Hydrocracking

In this process, gas oil or distillate is converted to lighter, higher octane blending components in the presence of hydrogen. Unlike an FCC unit, the process occurs over a fixed bed at high pressure. Because of the presence of hydrogen in the reactor, the product produced is saturated, with different blending properties than FCC product.

Table A-6 Hydrocracking Unit Capacities

Company	Location	Capacity (BPD)
LARGE		
ExxonMobil	Beaumont, TX	25,000 to 35,000
SMALL		
Wynnewood Refining	Wynnewood, OK	5,500

Naphtha Reforming

Many straight-run pipestill naphthas or condensates from natural gas liquid processing have low octane values due to the presence of paraffinic hydrocarbons. In order to increase the octane value and make the naphtha streams more suitable for blending, reformers are used. Reforming reactions usually occur at high temperatures over fixed-bed platinum catalysts. The product reformate is a branched, unsaturated hydrocarbon stream. Hydrogen is also produced in this reaction.

Table A-7 Naphtha Reforming Unit Capacities

Company	Location	Capacity (BPD)
LARGE		
BP	Whiting, IN	30,000 to 45,000
SMALL		
Somerset Refining	Somerset, KY	1,000

Desulfurization

Unless the crude slate is very sweet, most refinery gasoline and on-road distillate products require desulfurization to meet product specifications. This is a mature technology, using hydrogen and a fixed-bed catalyst to remove sulfur from the product stream.

Table A-8 Desulfurization Unit Capacities

Company	Location	Capacity (BPD)
LARGE		
ExxonMobil	Baytown and Beaumont	30,000 to 40,000
SMALL		
Montana Refining	Great Falls, MT	1,000

A.2 CLAUS SULFUR RECOVERY

A.2.1 Introduction

The Claus process is the principal process used for large scale (above ~25 TPD) sulfur recovery. The size of Claus plant used depends on the required degree of sulfur recovery, often dictated by air pollution regulations. The largest Claus plants are approaching 3000 TPD.

For smaller capacities (~1 TPD or less), a scavenging bed containing minerals, such as zinc oxide or liquid hexahydrotriazines, that react with sulfur is the predominant method of removal. At the intermediate scale (~1 to 20 TPD), regenerable sulfur recovery processes such as liquid reduction-oxidation (Redox), are more often employed. Examples of these processes include Stretford, LO-CAT II, and SulFerox, where chelated metal catalysts are used to convert H_2S directly to elemental sulfur. Economics make these plants attractive predominantly when the amount of sulfur removal is relatively low (under 20 TPD).

The need to process increasingly sour crude oil in refineries and increasing amounts of sub-quality gas containing H_2S will increase the need for future Claus plants. Technology improvements have mainly focused on tail gas treating processes to deal with ever tightening sulfur recovery requirements and means of increasing throughput in existing Claus plants. The basic modified Claus process is essentially unchanged since its widespread introduction in the 1950s.

A.2.2 Process Description

Claus plants are the major means of recovering sulfur from natural gas and refinery gas when processing requirements are ~25 TPD and above. The process involves removing the acid gas (CO_2 and H_2S) with an amine process or similar system upstream of the Claus unit, and then feeding the resulting acid gas from the amine stripper into a furnace (see Figure A-2). In the furnace, about 1/3 of the H_2S is converted to SO_2. The H_2S and SO_2 react to form elemental sulfur, with approximately 70% of the sulfur recovered in a condenser following the furnace. Multiple stages of catalyst beds with intermediate reheat are provided downstream of the Claus furnace to recover additional sulfur. Each stage reacts 2 moles of H_2S with one mole of SO_2 over a catalyst to produce elemental sulfur. A practical limit of 3 such stages has been found, since progressively less sulfur is removed in each stage. About 97% of the feed sulfur can be removed in this manner. If additional recovery is required, a Tail Gas Treatment Process (TGTU) is used.

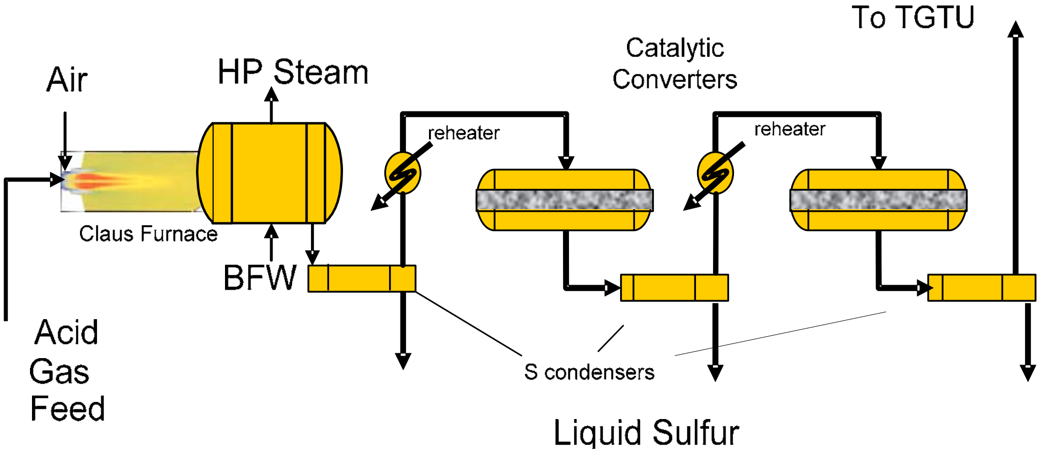

FIGURE A-2 TYPICAL CLAUS PLANT BFD[20]

A.2.3 Plant Capacity

Table A-9 below shows examples of commercial Claus units at both a large and small scale. Treatment capacity sizes range widely; as mentioned above, it is rare to find commercial units below 25 TPD. Examples of these types of facilities may have other factors that explain their existence, such as age or regulatory issues. It is unlikely that Claus technology would be chosen for a grass-roots facility today if the treatment capacity is under 25 TPD.

Other examples of Claus units built at sizes close to these can be provided, since the technology is commercially widespread.

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[20] Laurance Reid Gas Conditioning Conference, February 2000, courtesy of Charles Kimtantas, Bechtel

Table A-9 Claus Sulfur Recovery Capacities[21]		
Company	**Location**	**Capacity (TPD, S)**
LARGE		
Duke Energy Gas Transmission Co.	Pine River, B.C., Canada	2,000
SMALL		
San Joaquin Refining	Bakersfield CA, USA	3

A.3 NATURAL GAS REFORMING

A.3.1 Introduction

Natural gas reforming is the process of converting natural gas feedstocks to a synthesis gas consisting of CO, CO_2, and H_2 using steam as a reactant. The synthesis gas will also contain small amounts of unreacted methane and residual steam. This process is the first step in many industrial processes including methanol synthesis, ammonia manufacture, gas-to-liquids (GTL), and hydrogen production. The earliest steam methane reformers were introduced in the U.S. in the 1930s by Standard Oil of New Jersey at their Bayway and Baton Rouge refineries for hydrogen production from offgases[22].

In the U.S., ammonia and methanol plants, both users of SMR, have been closing as a result of competition with foreign plants with access to inexpensive natural gas. Overseas, large "Mega Methanol" plants in the Caribbean and in Asia are being built[23], along with large GTL plants in the Middle East (Qatar)[24]. Since these projects are all based on SMR, there is expected to be an increase in worldwide SMR capacity this decade. Capacity increases of SMR are also expected in refineries to produce hydrogen for processing the increasingly heavy, sour crudes being brought into petroleum refineries. In the long-term, if the US begins the transition to a "hydrogen economy", SMR will form the bridge providing hydrogen until advanced, non-hydrocarbon source hydrogen becomes feasible.

Several of the technologies discussed in other sections (Methanol, Ammonia and GTL) depend on production of synthesis gas from SMR. The syngas generation step is a very large portion of the total cost, on the order of 60% by GTI estimates. Therefore, in order to not unduly bias the modular cost study with what substantially amounts to a single technology, it would be advisable not to choose more than one technology that has SMR included in it.

A.3.2 Process Description

Natural gas reforming is the indirect heating of natural gas and steam in a furnace over a catalyst. The catalyst is typically nickel on alumina, and is usually packed within tubes suspended in the radiant section of a furnace. Hundreds of tubes are necessary, depending on the capacity of the reformer, each 25 to 40 feet in length and 3 to 5 inches in diameter. Sufficient steam must be provided to avoid carbon formation.

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[21] Oil and Gas Journal, June 24, 2002, Worldwide Sulfur Production chart

[22] Catalyst Handbook, Second Edition, Martyn V. Twigg, ed., Wolfe Publishing Ltd., 1989

[23] Gas-to-Liquid News, Methanol Watch, Vol VII, No. 10, October 2004

[24] "In Qatar, Oil Firms Make Huge Bet on Alternative Fuel", The Wall Street Journal, February 15, 2005

The reformate consists of CO, H_2 and CO_2, residual feed methane, and steam. High and low temperature shift reactors often follow SMR to adjust the H_2 to CO ratio of the product syngas. Following the radiant reformer section, there will be a convective zone where heat is recovered and steam for the process is produced. Since the catalyst is poisoned by sulfur, the feed is usually desulfurized to <1 ppm upstream of the reformer. Reformers are built by many companies including, ICI (Johnson Matthey/Synetix), Uhde, KBR.

The operating pressure is dependent on the desired composition downstream of the reformer, but economics usually dictate a range of 280 to 420 psig. The reformer furnace typically runs from 1400 to 1500 degrees F. Furnaces are typically fired with natural gas. Figure A-3 gives an overview of the process[25].

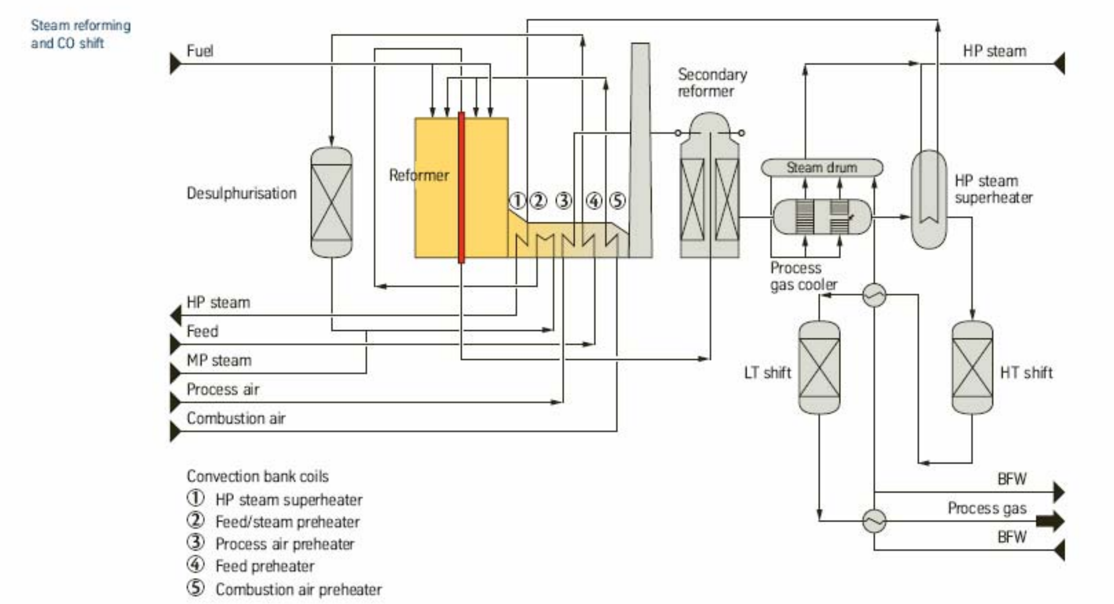

Figure A-3 Steam Methane Reformer Process Flow, Ammonia Plant

A.3.3 Plant Capacity

While technological improvements in SMR have been fairly limited, technical learnings, economies of scale, and downstream demands have promoted the development of ever-larger facilities. Considerable development has gone into mechanical issues such as tube metallurgy and schemes to attach the hot tubes to cold headers in order to prolong tube life and avoid shutdowns.

Table A-10 below shows examples of large and small scale SMR facilities. Large scale facilities tend to be associated with large chemical plants (ammonia or methanol production), while small scale reformers are more associated with hydrogen production. Besides today's commercial small scale plants, natural gas reformers are being built for vehicle hydrogen fueling stations and to produce hydrogen for fuel cells. GTI has built a prototype unit to produced 17,000 SCFD of hydrogen, although an expected commercial scale unit would be 2 to 2.5 times that capacity. This plant requires ~0.01 MMSCFD of natural gas feed.

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[25] Information from Uhde GmbH

Table A-10 SMR Plant Capacities[26]		
Company	**Location**	**Capacity (MMSCFD of natural gas)**
LARGE		
Burrup Fertilizers	Australia	72.6
SMALL		
Various (fuel cell applications)	various	0.1 to 0.01

A.3.4 Technology Advancements

KBR has developed the KRES (KBR Reforming Exchanger System) that integrates the offgas heat energy from the SMR to perform additional reforming in a specially designed reactor. There are two commercial KRES units, both in ammonia plants. The first was started up at Methanex's plant in Kitimat, BC, Canada in 1994 and has a capacity of 350 MTD. The second started up in 2003 at Liaohe's plant in Panjin, China and has a capacity of 1100 MTD. KRES cannot operate as a standalone system, however, but does lower the overall energy demand of syngas generation compared to SMR alone.

Developments in the area of autothermal reforming in lieu of conventional SMR are also in progress.

A.4 METHANOL FROM NATURAL GAS

A.4.1 Introduction

Methanol is an important industrial chemical and intermediate for other chemicals including silicones, refrigerants, adhesives, specialty plastics, coatings, textiles, and water-treatment chemicals. Methanol is produced from syngas that is reacted over a catalyst. The catalytic synthesis approach was introduced by BASF in 1923[27]. The primary source of syngas, similar to the situation for hydrogen production, ammonia synthesis, and GTL processes, is from reforming natural gas.

A.4.2 Process Description

The predominant commercial technology is the Low Pressure Methanol synthesis route pioneered by ICI (now Johnson Matthey/Synetix). This is a catalytic route using a copper/zinc oxide/alumina catalyst. The key design issue when dealing with methanol plants is controlling the heat release from the exothermic reactions. This is accomplished by a variety of means including placing the catalyst in tubes surrounded by water, interstage cooling, or quenching with feed in a multiple-bed reactor. Following this step, crude methanol is refined via distillation.

The manufacture of methanol requires synthesis gas that is usually produced from steam methane reforming (see Section 3 for a description of SMR). Following SMR, the synthesis gas is compressed to 700 – 1500 psig. The gas leaving the methanol synthesis reactor is cooled to

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[26] Information on the fertilizer plant from The Fertilizer Institute, http://www.eia.doe.gov/oiaf/aeo/conf/pdf/ vroomen.pdf

[27] Catalyst Handbook, Second Edition, Martyn V. Twigg, ed., Wolfe Publishing Ltd., 1989

condense the crude methanol product, with the remaining gas is recompressed, mixed with fresh feed, and returned to the methanol synthesis reactor. The crude methanol is then distilled to remove impurities. Figure A-4 gives a schematic representation of a typical methanol process[28].

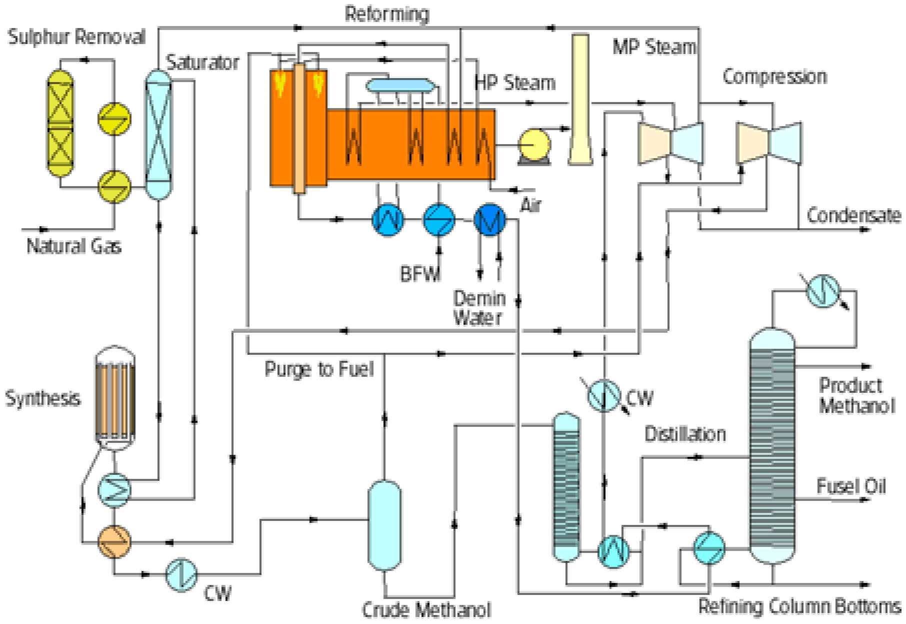

FIGURE A-4 LOW PRESSURE METHANOL PROCESS FLOW DIAGRAM

A.4.3 Plant Capacity

Methanol plants have been increasingly growing in size. The newest facilities, the so-called "Mega Methanol" plants, are exemplified by the 1.7 million tonnes per year Atlas Plant in Trinidad, which started up in mid-2004[29]. An example of large and small scale commercial facilities can be seen in Table A-11.

Table A-11 Methanol Plant Capacities

Company	Location	Capacity (metric tpa)
LARGE		
Methanex/BP	Pt. Lisas, Trinidad	1,700,000
SMALL		
Praxair	Geismar, LA	20,000

As in many other industries, the trend to larger sizes is to reduce the unit production costs. With the availability of low-cost natural gas feedstock in Trinidad, landed methanol prices are significantly less expensive than U.S. domestic sources using high-cost natural gas feed. As a result, numerous smaller, less efficient plants in the U.S. have closed. In 2003, a number of US plants ranging from 80,000 to 200,000 metric ton per year were shut down. With world

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[28] Information from Methanol Plant Technology – Synetix Low Pressure Methanol (LPM) Process Brochure, Synetix, 2003

[29] More information on this facility can be found at the Methanex website: http://www.methanex.com/ourcompany/locations_trinidad.html

methanol prices low, this trend is expected to continue with even larger plants. Estimates are that 4 million tons of annual methanol capacity will be shut down by 2008[30]. Larger plants with low-cost transportation options and adequate supplies of low-cost feedstocks will continue to be built as supply/demand balance allows. These will largely be Mega Methanol units.

A.5 AMMONIA PRODUCTION

A.5.1 Introduction

Ammonia is produced via the Haber-Bosch process by the reaction of nitrogen and hydrogen. The main source of hydrogen in the United States is from steam reforming of natural gas and partial oxidation of oil residues. The production of hydrogen by steam methane reforming is covered in Section 3. Nitrogen is introduced as compressed air into the secondary reformer. The hydrogen reacts with the nitrogen at high temperatures (750-1200ºF) and pressures (1500 to 3000 psig) over a catalyst, usually iron oxide with a potassium hydroxide promoter, although some new ruthenium-based catalysts are being suggested. The anhydrous ammonia is sold as product or used to produce a variety of fertilizers and other products, including urea, nitric acid, ammonium nitrate, and ammonium sulfate[31].

Figure A-5 shows the change in world-scale capacity over time. Prior to the 1970s, plants had much greater energy consumption rates, 10-12 Gcal/ton, compared to the current designs of less than 7 Gcal/ton.

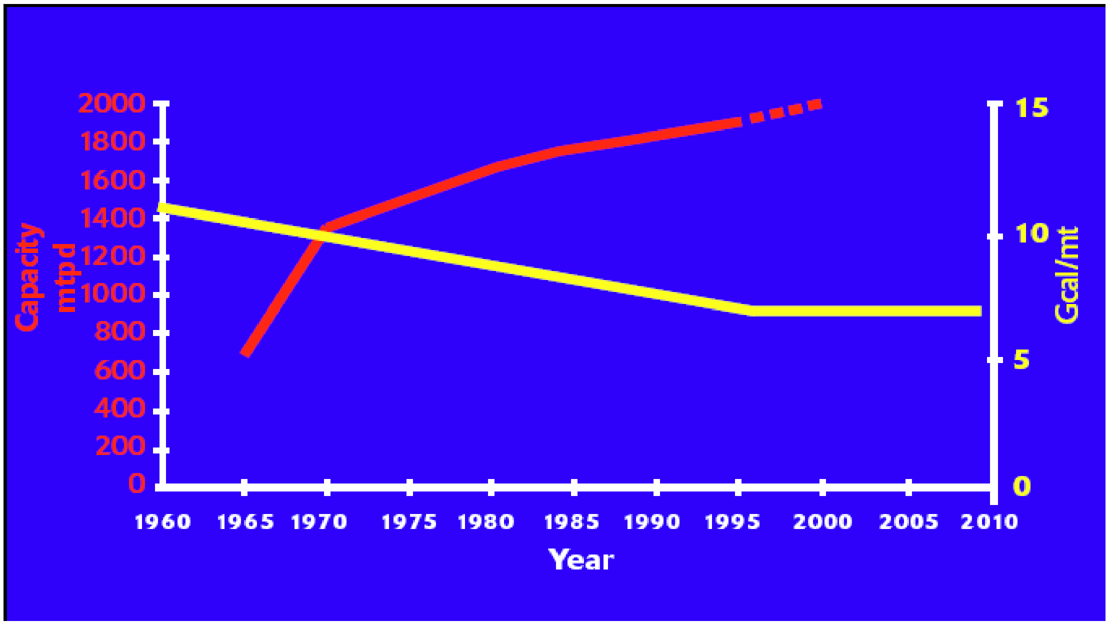

FIGURE A-5 AMMONIA PRODUCTION CAPACITY AND ENERGY CONSUMPTION[32] (1960-2010)

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[30] HPI Construction Boxscore, February 2005

[31] Process information from "Energy Use and Energy Intensity of the U.S. Chemical Industry", Worrell, et. al, Lawrence Berkeley National Laboratory, LBNL-44314, April 2000

[32] Information from Halliburton, available at http://www.halliburton.com/kbr/relatedInfo/chemicals /modamplt.pdf

A.5.2 Process Description

Ammonia is generally manufactured from natural gas using the steam reforming process as seen in Figure A-6. While other feedstocks and processes are used, this process represents the majority of the commercial designs seen today. There are several reaction stages and catalyst reactions key to the economic operation of modern ammonia production plants.

The first stage is purification where impurities, mainly sulfur compounds, are removed from the gas stream. Steam reforming is performed in two stages. In the primary stage, the endothermic reactions take place at pressures around 450 psig and temperatures of 1450°F or higher. This is followed by an exothermic secondary reformer, where air is added to the partially reformed gas stream.

The carbon monoxide in the gas leaving the secondary reformer is converted to carbon dioxide in the shift reactors. The carbon dioxide is then removed by scrubbing the gas stream via an amine process. Any residual carbon oxides are then converted back to methane before compression to ammonia synthesis pressure. The final reaction stage is ammonia synthesis where the hydrogen and nitrogen combine to form ammonia. This reaction stage takes place at high pressure (1500 to 3000 psig) and is highly exothermic. The removal of product ammonia is accomplished via mechanical refrigeration or absorption/distillation. Typically, refrigeration is more economic for high pressure operation, while absorption/distillation is favored for lower pressure operation.

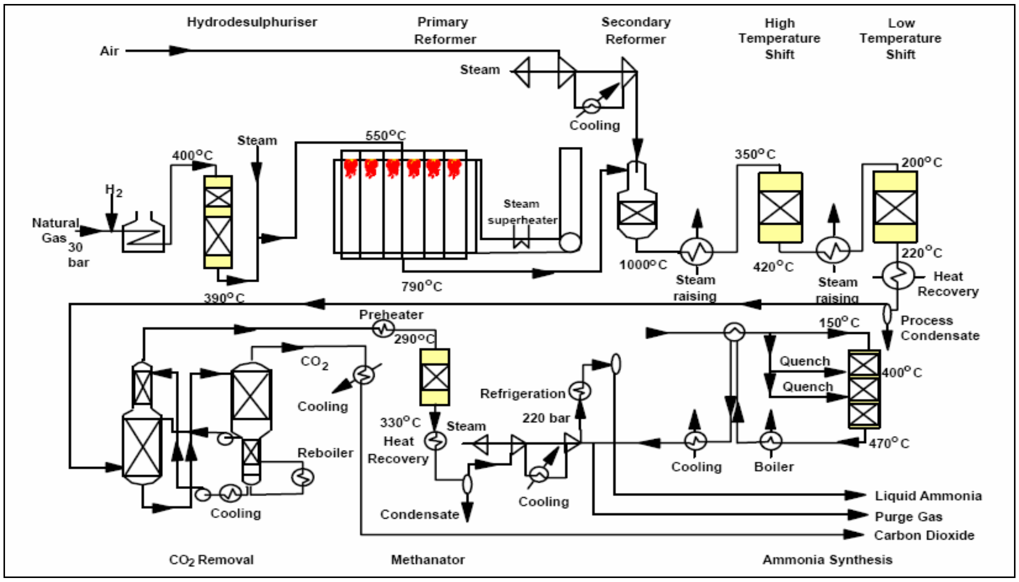

FIGURE A-6 AMMONIA PRODUCTION FLOWSHEET[33]

A.5.3 Plant Capacity

Major modern plants can easily produce 1,000 tonnes of ammonia per day, requiring almost 1 million m^3 of methane. Due to this high feedstock requirement and large plant economies of scale, production is moving to low cost hydrocarbon sources. Alternative sources of hydrogen

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[33] Information obtained from Synetix, Introduction to Ammonia Manufacturing, available at http://www.synetix.com/ammonia/pdfs/ammonia.pdf

are being sought, including coal via gasification and water via electrolysis using an alternative energy source, such as solar. Many ammonia plants in the United States are currently shut down because the high cost of feedstock results in production costs greater than the market value of the produced ammonia[34].

Table A-12 Ammonia Plant Capacities		
Company	**Location**	**Capacity (TPD)**
LARGE		
Burrup Fertilizers	Australia	2,200
SMALL		
Shoreline Chemical	Gordon, GA	31

A.6 GAS-TO-LIQUIDS

A.6.1 Introduction and Summary

GTL or Gas-to-Liquids, represents a number of processes which have been proposed to monetize stranded natural gas by manufacturing mainly low sulfur diesel and side products like waxes and lube oils from low-cost natural gas. Production technology is essentially synthesis gas manufacture by reforming, followed by proprietary synthesis technology based on Fischer-Tropsch catalytic chemistry.

Only two commercial plants exist: the Shell Bintulu facility in Malaysia, and Sasol's commercial technologies in South Africa. All others facilities are either pilot units or are currently in the design stages. A number of process technologies exist for conversion of natural gas to Fischer-Tropsch liquids. All GTL technology is highly proprietary, making it difficult to provide accurate process and cost information. With the potential for wider commercial emergence of the technology in the future, the GTL market is highly competitive unlike semi-public domain technology like SMR and methanol production. While small scale GTL test facilities exist, costs for pilot plants are roughly 3 times as much as they would be for a scaled-down unit of same size based on commercial technology[35]. Costs for pilot scale facilities can also be deceptively low because, for example, they may not include the product upgrading step or do not require a commercial oxygen facility for a unit of that size.

Because of the difficulties in developing accurate cost and design information for GTL plants with published data, it is not recommended that they be considered for future design work.

A.7 NATURAL GAS LIQUIDS EXPANDERS

A.7.1 Introduction

Natural gas that is extracted from terrestrial sources consists predominantly of methane, and, to a lesser extent, a quantity of Natural Gas Liquids (NGL). These are typically ethane and propane, but also contain some butane and lesser amounts of higher carbon content, aliphatic

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[34] US Geological Survey (USGS), available at http://pubs.usgs.gov/of/2004/1290/).

[35] Based on GTI estimates and Merrow, Phillips, and Myers, Understanding Cost Growth and Performance Shortfalls in Pioneer Process Plants, The RAND Corporation, 1981

hydrocarbons. Table A-13 below shows compositions for a range of potential feedstocks to a NGL plant.

Table A-13 Typical NGL Feedstock Compositions

	Classification			
	Super-Rich	Rich	Lean	Super-Lean
C1 (mole%)	82.00	87.00	92.00	96.30
C2	9.70	7.90	4.80	2.20
C3	4.50	3.30	1.80	0.80
IC4	0.50	0.46	0.49	0.20
NC4	1.80	1.13	0.87	0.50
IC5	0.50	0.00	0.00	0.00
NC5	0.49	0.21	0.04	0.00
C6+	0.51	0.00	0.00	0.00
BTU/scf	1251.81	1161.77	1105.35	1054.51

If the quantities of these materials are low, they slightly enhance the heating value of the natural gas and may be left in the gas. If the amount of such materials is such that there may be condensation in the transmission pipeline, or there could be an impact on the suitability of the gas for various applications (usually measured in terms of the heating value and/or the related Wobbe number), the pipelines may not purchase the gas from producers without prior treatment to remove the heavy hydrocarbons.

In addition, a large industry has grown to utilize the ethane/propane and heavier hydrocarbons being removed from the natural gas as petrochemical feedstocks. Extraction of NGLs from natural gas can be either profitable or an unavoidable cost of production of natural gas for pipeline transport, depending on the relative price of the hydrocarbons on a heating value and chemical feedstock basis.

A.7.2 Process Description

Early facilities used circulating heavy hydrocarbon liquids to absorb the lighter components in lean oil absorber plants. In the 1960s, advances in turbomachinery led to newer cryogenic processes being introduced. These turboexpander processes have become the dominant technology.

Technology has evolved mainly to enable the expander plants to have reasonable recoveries and economics when rejecting ethane (leaving it in the gas) while recovering 90% or more of the propane and heavier components. This is necessary to enable such plants to remain in operation when economics favor leaving the ethane in the pipeline gas. There are a number of patented schemes that improve upon the original turboexpander design in order to enable this mode of operation. Many companies offer NGL expander plants including Fluor, Ortloff, IPSI LLC, ProQuip, and ABB Randall.

Figure A-7 shows the schematic of the NEPTUNE II NGL extraction plant built and operated by Enterprise at St. Mary Parish, LA. This plant uses technology from IPSI LLC, a subsidiary of Bechtel, and has a gas processing capacity of 300 MMSCFD. The NGL content of the inlet gas is about 4.0 gallons per MSCF. Approximately 88% of C2, 99% C3 and 100% of C4+ are recovered as NGLs. The shrinkage, or loss of feed natural gas resulting from the processing, is about 11-12% of the inlet gas throughput. This plant was commissioned in early 2004.

The process for the removal of NGL from natural gas is essentially a mechanical refrigeration scheme. There are no chemical reactions taking place. Natural gas is fed at a pressure above 400 psi, is exchanged with exiting cold gas from the top of the demethanizer column, then is sent to a cold separator to drop out condensed liquids. The condensed liquids are further cooled and sent to the demethanizer column as cold reflux. The gas exiting the cold separator is expanded in the turboexpander that cools and condenses significant amounts of liquid for column reflux. The demethanizer distills the cold liquid and gas feeds to the column. The overhead gas is highly concentrated in methane with some ethane if desired. The column bottoms will be primarily ethane and propane. Depending on the gas composition and the NGL product specifications, it may be necessary in some plants to further distill the liquid to remove butane and heavier components.

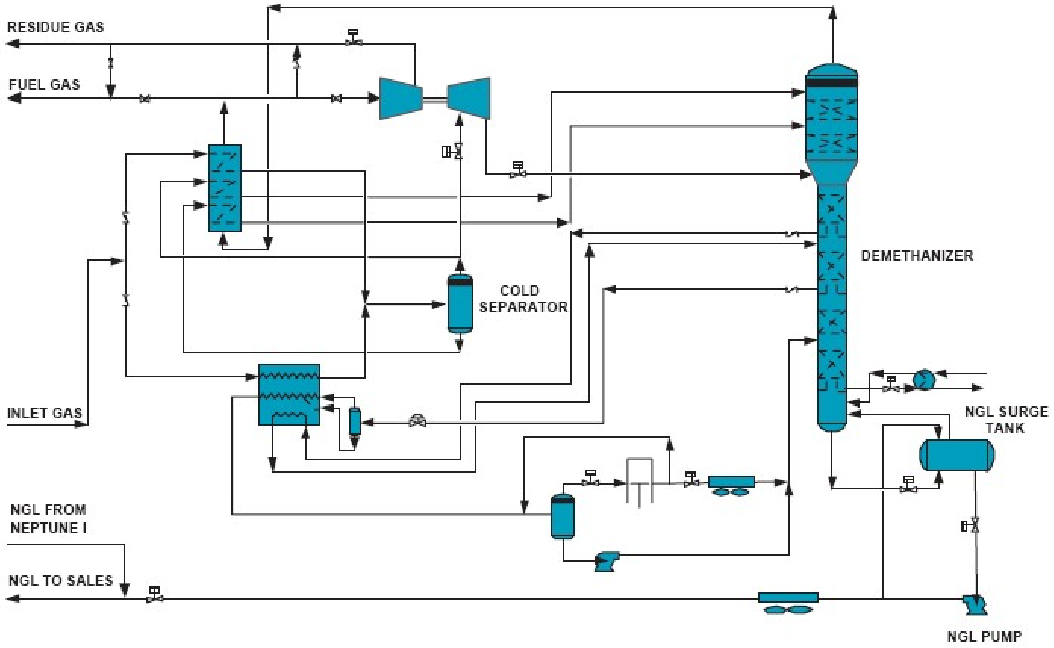

FIGURE A-7 SCHEMATIC FOR NEPTUNE II TURBOEXPANDER PLANT[36]

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[36] Oil and Gas Journal, July 21, 2003

A.7.3 Plant Capacity

Plants are constructed of a size commensurate with the amount of NGL expected in the natural gas feed. Natural gas contains NGLs at varying concentrations, and different producing regions may have varying amounts of natural gas being produced depending on the age of the field and the number of wells being drilled. A fairly wide range of expander plants sizes are currently being built. This is unlike other process technologies mentioned during this study, where economies of scale have led to increasingly larger facilities.

The number of expander plants in the U.S. is on the order of 300 – 400. The largest facilities have capacities on the order of 2,500 MMSCFD, with multiple trains of processing. Smaller, skid-mounted plants may be as small as 10 to 50 MMSCFD. Table A-14 gives examples of typical large and small scale NGL Expander facilities.

Table A-14 NGL Turboexpander Plant Capacities

Company	Location	Capacity (MMSCFD)
LARGE		
Enterprise	St. Mary Parish, LA	300 - 350
SMALL		
Aqila	Sommerville, TX	25

A.8 CRYOGENIC OXYGEN PLANTS

A.8.1 Introduction

Oxygen was first extracted from the atmosphere by a chemical process. This was superseded in the 1920s by the cryogenic process involving low temperature liquefaction and distillation of air[37]. Over the years, the sizes of the oxygen plants have grown from 100 TPD to almost 4,000 TPD. The cryogenic air separation process is by far the most widely used. However, non-cryogenic techniques first developed during the 1970s – adsorption technologies – are becoming increasingly significant for smaller or lower purity applications.

The choice of oxygen production technologies depends on several factors: the volume required, the purity required, the customer location, and the oxygen usage pattern. For the large tonnage oxygen customers (over roughly 300 TPD), a cryogenic plant is usually more economic. For the customers requiring lower oxygen flow rates, adsorption or delivered oxygen is usually preferred. Cryogenic plants cannot compete in the low tonnage area primarily because of the difficulty of scaling down the distillation column and the poor efficiencies of the small compressors. Non-cryogenic plants, on the other hand, are limited in their sizes because of the constraints of the adsorber diameters.

The major oxygen suppliers are Praxair, Air Product, BOC Gases (UK), Air Liquide (France), and Linde (Germany). Hundreds of cryogenic and non-cryogenic plants have been installed worldwide.

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[37] Information on technology history from BOC gases, available at http://www.boc.com/capability/gases /air_separation.asp

A.8.2 Process Description

In the cryogenic process, filtered air is first compressed and cooled by the main air compressor and is sent to the adsorbers, where water, carbon dioxide and any hydrocarbons are removed from the air stream. The clean dry air is sent to a series of heat exchange and distillation steps to separate the major constituents: oxygen, nitrogen, and argon.

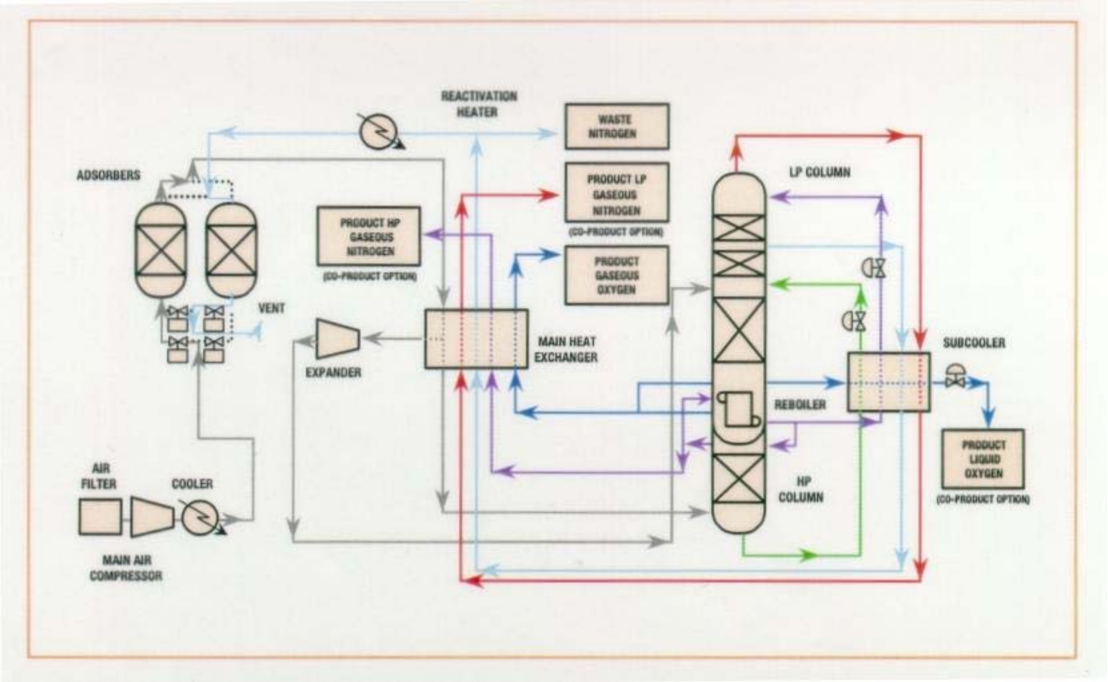

FIGURE A-8 GENERIC CRYOGENIC OXYGEN PLANT FLOW DIAGRAM[38]

A.8.3 PLANT CAPACITY

As with many other process technologies discussed earlier, cryogenic oxygen plants do have an economy of scale that has led to larger and larger plants in recent years. Most small facilities either are older, or have specialty niche applications that justify their use. Table A-15 below shows examples of small and large scale cryogenic oxygen facilities[39].

Table A-15 Cryogenic Oxygen Plant Capacities		
Company	**Location**	**Capacity LOX (TPD)**
LARGE		
Air Liquide	Secunda, South Africa	3,550
SMALL		
BOC	Lahore, Pakistan	30

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[38] From Air Products, available at http://www.airproducts.com/NR/rdonlyres/C13A9654-7BB1-4337-A4D9-94F64546A672/0/APACK.PDF

[39] Air Liquide information from Engineering News, available at http://www.engineeringnews.co.za/eng /features/sasol. BOC information from "First Move of Aspen Plant Ends in Success", FrostByte – Newsletter of Cryogenic Industries, Vol. 10, No. 4, Spring 2000

A.9 ETHYLENE AND OTHER OLEFINS PRODUCTION

A.9.1 Introduction

Ethylene is used primarily as an intermediate in the manufacturing of plastics and other chemicals. It is polymerized to form polyethylene, chlorinated to form the precursors for polyvinyl chloride, or reacted with benzene in the manufacturing of polystyrene. Other chemical uses include the production of ethylene oxide, ethanol, polyvinyl acetate, and ethylene glycol. It also has a small market as a plant hormone for ripening picked fruit.

The primary route to the production of ethylene is by thermal cracking of hydrocarbons. Other paths, including catalytic pyrolysis, oxy-dehydrogenation of ethane, oxidative coupling of methane, and dehydration of ethanol have limited or no commercial interest. In thermal cracking, gaseous or light hydrocarbons are briefly heated to 1500-2000ºF, initiating a free radical reaction, breaking down the heavier hydrocarbon chains into smaller ones, and unsaturating previously saturated hydrocarbons. The end product is a complex mixture of ethylene, propylene, butadiene, benzene, fuel oil, and other products, requiring extensive separations and purification steps.

A.9.2 Process Description

The heart of the ethylene plant is the cracking furnaces. These are tubular reactors in which the pyrolysis reactions take place. Before 1960s, the cracking tubes were horizontal rows in a radiant chamber. This gave low ethylene yields, leading to more modern designs using vertical tube rows to provide superior mechanical performance and higher capacity. Steam is added to the feed to prevent coking and to minimize side reactions. The product mix ranges from hydrogen to tars. This requires extensive separations to meet product purity specifications. This is performed by compression, drying, fractionation, absorption, adsorption and catalytic conversions.

A simple flowsheet for an ethylene production process is shown in Figure A-9[40]. Figure A-9 shows a plant with a relatively light feedstock mix, requiring little separation of hydrocarbons heavier than propane. The complexity of the flowsheet will depend on the feedstock and product mix desired.

Feedstock enters the plant and is fed to the pyrolysis heaters after preliminary heat exchange. Here the feedstock is broken down to lighter hydrocarbons. The remainder of the plant is separations and purification. The gasoline and fuel oil fractions are recovered first through quench and stripping that removes the heaviest fractions. The vapor fraction is then compressed to over 500 psig for additional purification via fractionation and cold box steps. Acid gas removal may be performed between compression stages[41].

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[40] Information from Halliburton, available at http://www.halliburton.com/kbr/relatedInfo/chemicals/ ethylene.pdf

[41] Technical information from Encyclopedia of Chemical Technology, Fourth Edition, Kirk-Othmer, John Wiley & Sons, NY, 1994

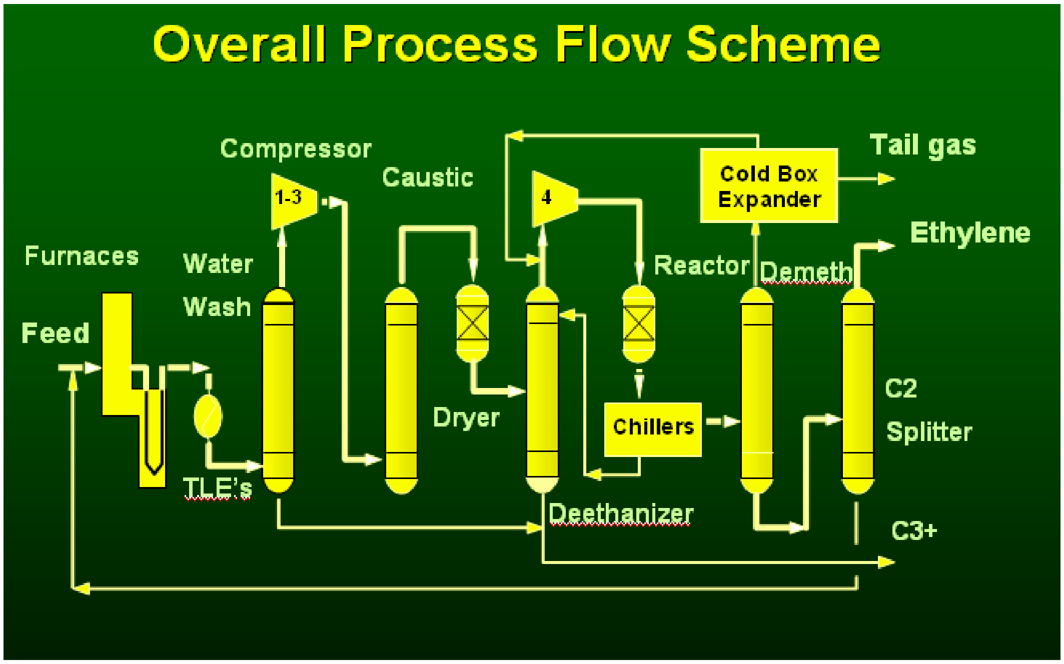

FIGURE A-9 KBR SCORE™ (SELECTIVE CRACKING OPTIMUM RECOVERY) ETHYLENE TECHNOLOGY

A.9.3 Plant Capacity

The capacity changes over time are shown in Figure A-10[42]. The solid line is an estimate of the maximum plant capacity able to be built during that year. Through the 1960s and 1970s, steam cracking focused on gaining economies of scale by dramatically increasing the size of each plant. The increase was facilitated by the evolution of compression equipment to highly reliable, large centrifugal compressors, improvements in furnace design, a better understanding of light ends fouling, and advances in process control. Also significant were improved computerized simulations that made design of today's modern recovery equipment possible.

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[42] From R. Orriss, "Major Investment Opportunities in the 21st Century, The Changing World of the Global Ethylene Business," Global Ethylene Conference, Saudi Arabia, 2004.

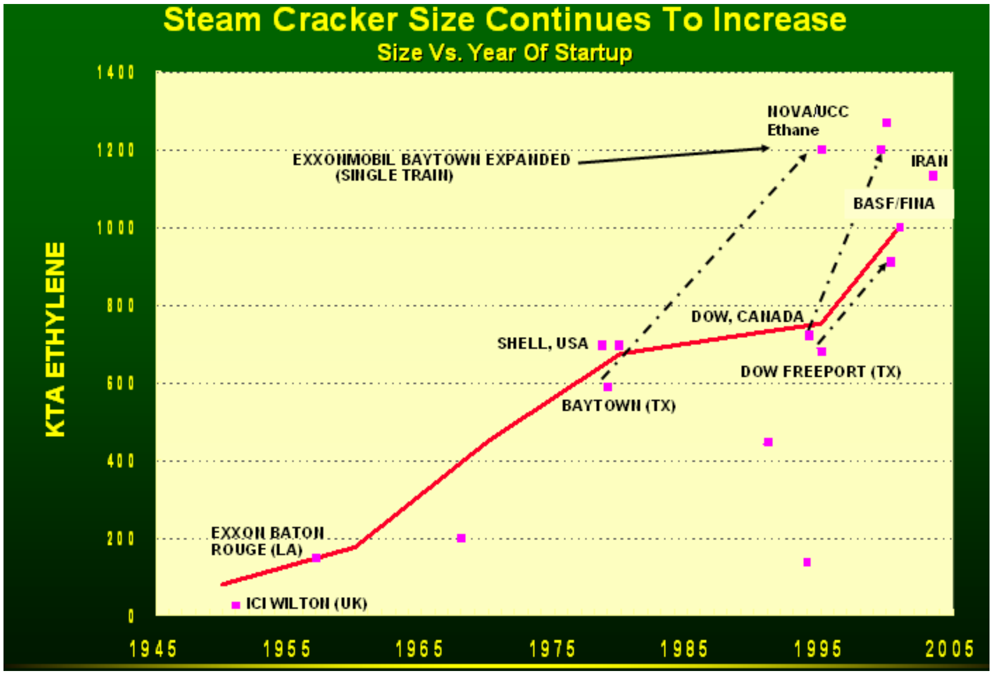

FIGURE A-10 ETHYLENE PLANT SIZE PROGRESSION, 1945-2005

The drive to build larger single train steam crackers slowed after about 1980. This was not caused by technological barriers, but by the difficulty of marketing large quantities of product and acquiring enough feedstock. Focus shifted to debottlenecking existing units. The ExxonMobil plant at Baytown, Texas, is currently the largest single train in the world at about 1,200,000 tons/yr. As the dotted line shows, this plant was built in 1979 with an original capacity of 590,000 tons/yr. Through several debottleneck programs, the capacity has been doubled at far below grassroots cost. With improved heat transfer and mass flow inside the furnace tubes, better process models, extremely short times in the reaction zone, and rapid quench systems, the world-class plants have the capacity to expand from 600,000-1,000,000 tons/year to over 1,500,000 tons/y.

Table A-16 shows examples of large and small ethylene plants currently in operation. Small scale facilities usually have some sort of feedstock advantage that justifies their continued operation. For example, niche facilities with low-cost liquids that have superior economics for ethylene production over blending to gasoline or other refinery processes can economically exist.

Table A-16 Ethylene Plant Capacities[43]

Company	Location	Capacity (TPD)
LARGE		
ExxonMobil	Baytown, TX	3,500
SMALL		
Javelina	Corpus Christi, TX	108

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[43] Information on plant sizes from team personal experience, Oil & Gas Journal, March 22, 2004, and Chemical Markets Reporter, October 6, 2003

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This deliverable is the Final Report for Task 1, Cost Estimates of Small Modular Systems, as part of NREL Award ACO-5-44027, “Equipment Design and Cost Estimation for Small Modular Biomass Systems, Synthesis Gas Cleanup and Oxygen Separation Equipment”. Subtask 1.1 looked into processes and technologies that have been commercially built at both large and small scales, with three technologies, Fluidized Catalytic Cracking (FCC) of refinery gas oil, Steam Methane Reforming (SMR) of Natural Gas, and Natural Gas Liquids (NGL) Expanders, chosen for further investigation. These technologies were chosen due to their applicability relative to other technologies being considered by NREL for future commercial applications, such as indirect gasification and fluidized bed tar cracking. Research in this subject is driven by an interest in the impact that scaling has on the cost and major process unit designs for commercial technologies. Conclusions from the evaluations performed could be applied to other technologies being considered for modular or skid-mounted applications.

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biomass; Nexant; small modular; synthesis gas; syngas; fluidized catalytic cracking; steam methane reforming; natural gas liquids expanders; scaling; National Renewable Energy Laboratory; NREL

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