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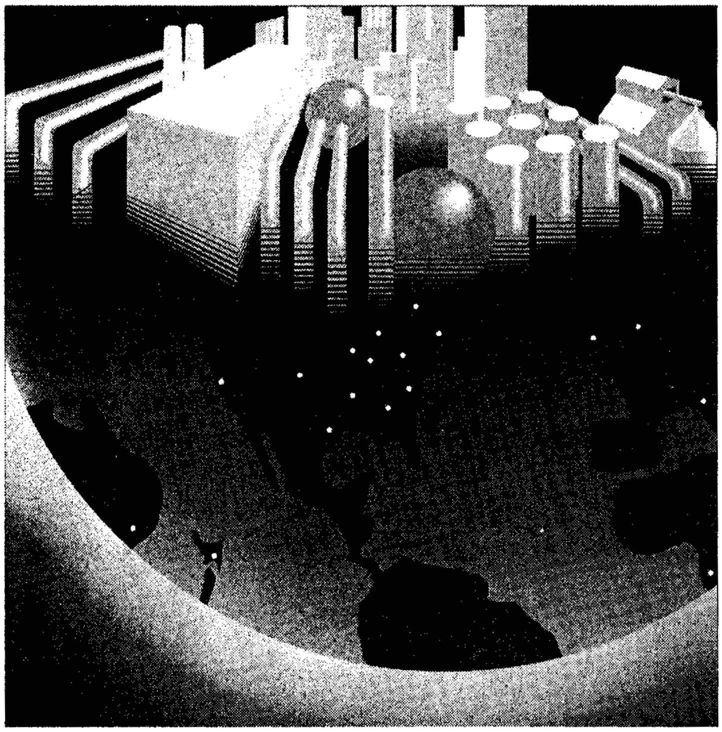

Volume 6: Case Study - Pennzoil's Atlas Site

CEC Report CR-105237

Prepared by
Linnhoff March, Inc.
9800 Richmond Avenue, Suite 560
Houston, TX 77042
May 1995

CAP 0519

REPORT SUMMARY

PINCH TECHNOLOGY/PROCESS OPTIMIZATION
Volume 6: Case Study - Pennzoil's Atlas Site

The petroleum refining case study of Pennzoil's Atlas refinery at Shreveport, LA demonstrates how process integration or pinch technology can identify practical and cost-effective ways to substantially reduce energy costs. Suggested cost-saving measures include steam and power system improvements and optimum heat exchanger network design. Energy savings in excess of $1,640,000 were identified with an overall payback of 0.72 years.

INTEREST CATEGORIES
Industrial
Demand-side Planning

KEYWORDS
Pinch technology
Heat recovery
Energy efficiency
Heat pumps
End use
Industry

BACKGROUND Improved industrial process efficiency is of great importance to electric utilities. It enhances customer competitiveness and profitability, thereby fostering load retention and strategic load growth. By understanding the energy use patterns and options at an industrial site, the utility can work together with its customer to define mutually beneficial investment and operating options. The technique of choice is pinch analysis, an innovative and effective method for analyzing industrial sites. Since 1988, EPRI and member utilities have cosponsored over twenty such studies around the country in various industries, with a high degree of success.

OBJECTIVES To identify opportunities for energy savings using pinch technology; to develop technically and economically viable projects to achieve these targets.

APPROACH Project team was formed consisting of consultants, plant and electric utility representatives. The team visited the plant to define and collect process, utility and economic data. The consultants developed appropriate material, heat and steam balances, and using pinch technology, characterized each refinery's heating and cooling needs. After quantifying the scope of potential improvements, the site was screened for specific projects based on processing changes, heat recovery and heat pump applications.

RESULTS The study indicates that substantial operating cost savings could be achieved using conventional technologies and investment payback criteria(typically two years or less). Projects were identified to reduce energy cost through improvements in the steam and power system and through the proper design of the heat exchanger network.

EPRI PERSPECTIVE The study shows that process integration or pinch technology is an effective tool to improve industrial energy efficiency. Utilities can use pinch methodology to promote load stability in their service territories. Information on energy use options and interactions and their sensitivity to economic factors can also be used to foster successful demand-side management programs.

EPRI has published additional case studies of pinch technology on a variety of industries. These are documented in reports TR-101147, Volumes 1 through 5. Other related work is documented in EPRI reports EM-6057, CU-6334, CU-6775, CR105238, and CR105239. An EPRI brochure on pinch technology is numbered BR-102466.

PROJECTS
RP3879-3
EPRI Project Manager: K. R. Amarnath
Customer Systems Division
Contractor: Linnhoff March, Inc

For further information on EPRI research programs, call EPRI Technical Information Specialists (415) 855-2411.

PINCH TECHNOLOGY/PROCESS OPTIMIZATION

Volume 6: Case Study - PENNZOIL'S ATLAS SITE

Final Report, June 1994

Prepared by
Linnhoff March, Inc.
9800 Richmond Avenue, Suite 560
Houston, TX 77042

Prepared for
Electric Power Research Institute
3412 Hillview Avenue
Palo Alto, CA 94304

For technical information, contact:

EPRI Chemicals & Petroleum Office
1775 St. James Place, Suite 105
Houston, Texas 77056
(713) 963-9307
Fax: (713) 963-8341

For ordering information, contact:

EPRIAMP Customer Assistance Center (ECAC)
1-800-4320-AMP

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ABSTRACT

Pennzoil is expanding its Atlas refinery at Shreveport, LA. EPRI and Southwestern Electric Power Company (SWEPCO) provided assistance in this project by partially funding an energy efficiency study for the expansion. This was carried out by Linnhoff March, Inc.. In the study, pinch analysis was used to evaluate the process design and identify ways to reduce energy consumption.

Pinch analysis uses data on temperature, enthalpy, heat transfer coefficients, utility costs, and equipment costs to develop "targets" for energy and capital. Based on these targets, and the fundamental insights that the technique gives into the heat flows in the process, options are developed for saving energy economically. This procedure has been used to find savings in a wide range of industrial processes. In the present study, improvements were identified both in the individual process units and in the plant utility systems that serve them. Energy savings in excess of $1,640,000 were identified, with an overall payback period of 0.72 years.

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EXECUTIVE SUMMARY

Pennzoil is adding gasoline production units at its Atlas Refinery in Shreveport, Louisiana. The main expansion units are a Resid Catalytic Cracker (RCC) with a gas plant, and a sulfuric acid alkylation unit. Linnhoff March was contracted to conduct an energy optimization study for the plant expansion using pinch analysis. The Electrical Power Research Institute (EPRI) and Southwestern Electrical Power Company (SWEPCO) provided financial support for the study. The work was carried out in March 1994 while the project was undergoing preliminary engineering in Jacob Engineering's office in Houston, Texas. The RCC generates approximately 124,000 lb/hr of steam, and its associated gas plant has a steam demand of 14.97MMBtu/hr based on preliminary design data. Linnhoff March determined that the steam demand could not be reduced economically. However, an opportunity was identified to recover 8.94 MMBtu/hr into Deaerator Feed Water. The preliminary alkylation unit design has a steam load of 79.7 MMBtu/hr. Through process modifications Linnhoff March has recommended projects to reduce this by 13.5 MMBtu/hr. These projects save $350K and $530K/yr in the RCC and alkylation unit, respectively. In addition, several projects were identified for improving this site's steam system. The recommended projects are tabulated on the next page.

Project 1 is simply housekeeping. Additional (unquantified) housekeeping savings should be possible, especially through steam leak repair . Project 2 resulted from a review of a steam model for the existing plant and new units. Project 3 is a project for the RCC/ gas plant. Projects 4,5,and 6 are for the alkylation unit. The list notes only the recommended projects. Other projects were found, but were judged not recommendable for either financial or process reasons.

Description	**Savings $'000/yr**	**Capital Cost $'000**	**Payback yr**
1. Blowdown Exchanger	150	n/a	n/a
2. 50 # Steam Turbine (Air Blower)	475	771	1.62
3. RCC +LCO/DFW Preheat	383	67	0.2
4. DeC3- Increased Feed Heat	128	131	1.02
5. Uses for DeC4 Bottoms (**either a or b**)			
5a. City Water for Wash*	137	29	0.21
5b. DIB Feed Preheat	132	45	0.34
6. Intermediate Reboiler on DIB Column (**either a or b**)			
6a. Raise DeC4 Pressure*	367	192	0.52
6b. MVR on DeC4 Overheads	436	1,400	3.2
Total including * options	1640	1,190	0.72

LIST OF ABBREVIATIONS

BFW	Boiler Feed Water
BPSD	Barrels per Stream Day
Cat Cooler	Catalyst Cooler
CO Boiler	Carbon Monoxide Boiler
DeC3	Depropanizer Column
DeC4	Debutanizer Column
DIB	DeisoButanizer Column
DFW	Deaerator Feed Water
DOE	Department of Energy
EPRI	Electric Power Research Institute
H+MB	Heat and Material Balance
HEN	Heat Exchanger Network
LCO	Light Cycle Oil
LMI	Linnhoff March Incorporated
LPG	Liquefied Petroleum Gas
MMBtu/hr	Millions of British Thermal Units per Hour
MVR	Mechanical Vapor Recompression
PFD	Process Flow Diagram
psig	Pounds per Square Inch Gauge
RCC	Resid Catalytic Cracker
SWEPCO	Southwestern Electric Power Company
VGO	Vacuum Gas Oil
ΔT_{min}	Minimum Temperature Difference

1

INTRODUCTION

Background and Objectives

Pennzoil is expanding its Atlas Refining facility in Shreveport, Louisiana. The main plant additions are a Resid Catalytic Cracking (RCC) unit, the Gas Plant associated with the RCC, and an Alkylation unit. Jacobs Engineering Group is carrying out preliminary design work for the new units. The expansion will increase the electric power requirements at the Atlas refinery, and Southwestern Electric Power Company (SWEPCO) will be providing additional electric service facilities.

Pennzoil, in conjunction with SWEPCO and the Electrical Power Research Institute (EPRI), requested Linnhoff March Incorporated (LMI) to perform an Industrial Efficiency Optimization using pinch technology for the plant expansion. The study, while having a component of energy reduction in its scope, also addressed broader issues relating to the site steam/power balance, selection of drivers for rotating equipment, and possible heat pump applications.

Scope of Work

The study was carried out in parallel with the design effort at Jacobs Engineering in Houston, Texas. Its purpose was to produce targets for energy consumption, schemes to improve process efficiency, approximate payout times for the various options, and an overview of the site's steam balance with the addition of the new units.

2

STUDY BASIS

The principal areas studied were the RCC unit (which will process 4480 BPSD of VGO and 4000 BPSD of resid); the accompanying gas plant; the new alkylation plant (which will produce 3900 BPSD of true alkylate); and the existing site steam system. In addition, steam demands from the other units to be added; sour water stripper, amine unit, Butamer unit, and sulfur plant; were included in the analysis to assess their impact on the overall site steam balance. The RCC/gas plant and the alkylation unit were examined as separate entities. However, the steam system was considered to be common to all areas. For the purpose of economic analysis, the plant was assumed to operate 8300 hrs/yr and a simple payback of 4 years or less was required for any capital project.

Process

RCC/Gas Plant

Resid catalytic cracking units produce components for gasoline blending by breaking long chained "bottom of the barrel" hydrocarbons. An important aspect of the process is the production of coke on the cracking catalyst from the chemical reactions. This coke leads to catalyst deactivation and therefore must be constantly removed by a regeneration step. Catalyst reactivation is a highly exothermic reaction and the heat is used to produce steam.

The outlet stream from the fluidized bed reactor contains a wide range of hydrocarbons which need to be separated in a main fractionator. The products from this distillation column are overheads, LCO, and slurry oil. These products go to gas treating, diesel dewaxing, and fuel oil storage, respectively. The gas treating unit basically consists of an absorber, stripper, depropanizer, and depentanizer. The absorber removes propane and heavier fractions from the feed. These are separated in the fractionation columns to produce C3/C4/C5 and gasoline. These are sent to the alkylation unit and gasoline blending, respectively. The absorber overhead is fuel gas going to the fuel gas system. Figure 2.1 contains a simple block flow diagram of the process, and Table 2.1 is the stream data used for pinch analysis. The Cat Cooler and CO Boiler have intentionally been omitted from the stream data, as one of the constraints of this study was that the heat recovered from these units can only be used for steam generation, and not for integration with other process units.

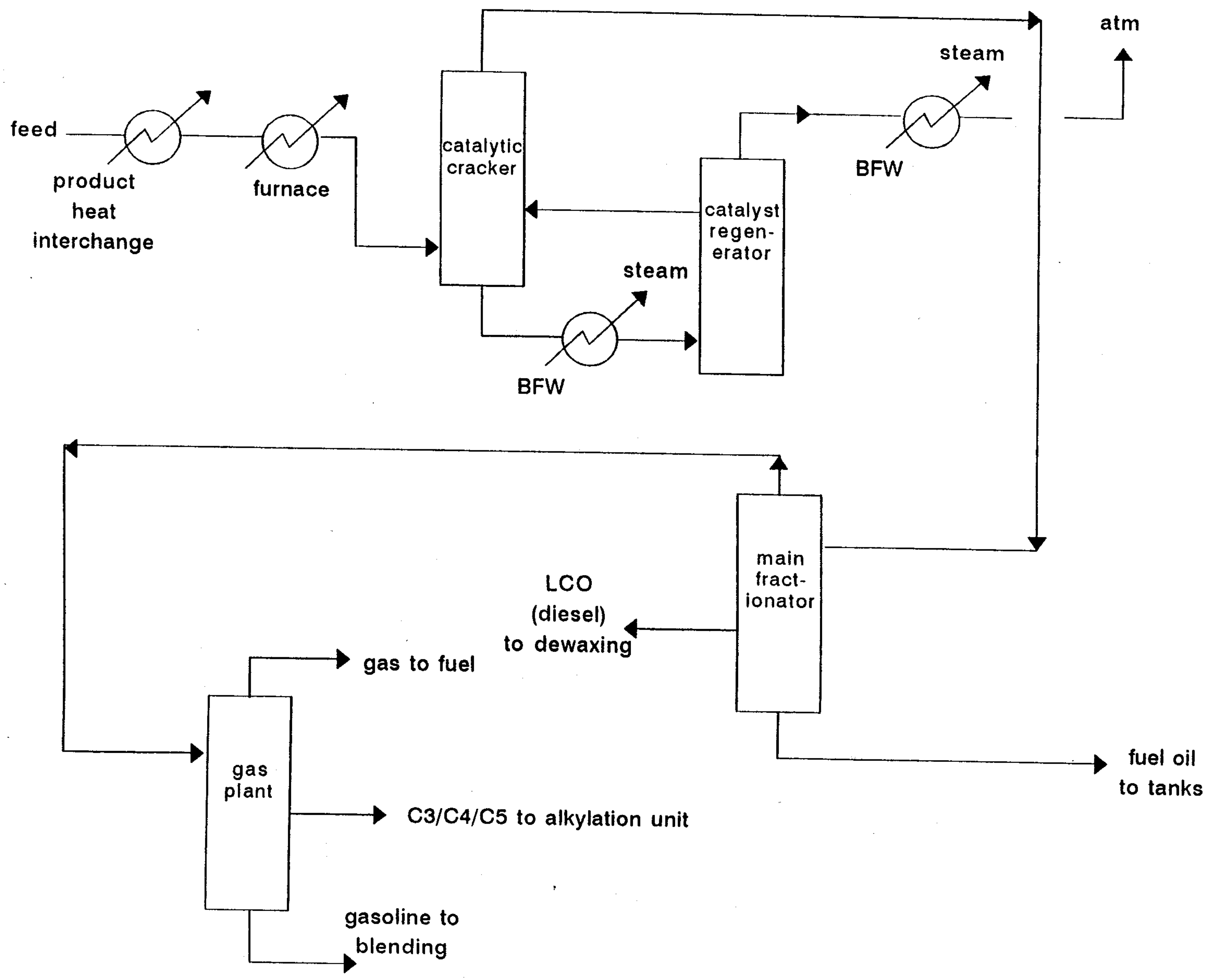

Figure 2.1
RCC Plant Flow Scheme

Table 2.1
Atlas RCC/Gas Plant
Stream Data

STREAM NAME	SUPPLY TEMP. [F]	TARGET TEMP [F]	ENTHALPY CHANGE (2)	MCP (1)	HTC (3)
SLURRY P/A	659	485	14.6	0.151	9.70E-05
CLARIFIED OIL PROD	659	175	2.2	8.18E-03	9.70E-05
HCO P/A	578	430	16.6	0.202	0.00012
LCO P/A	428	290	10.1	0.132	9.80E-05
LCO P/A TO SPONGE	290	185	2.3	3.93E-02	9.80E-05
	185	105	1.5	3.47E-02	9.80E-05
LCO PROD	417	100	3.5	1.96E-02	9.80E-05
C3 BOTTOMS	311	260	4	0.142	0.00017
RCC PROD	353	100	11.1	7.89E-02	0.00017
MAIN FRAC OVHD	192	100	26.7	0.523	0.00011
WGC INTERSTAGE	130	100	3.5	0.212	0.00011
WGC DISCHRGE	176	100	11.9	0.283	0.00011
ABS TOP P/A	118	100	0.7	7.26E-02	0.00017
ABS BOTT P/A	113	100	0.5	7.27E-02	0.00017
DE C3 OVHD	112	106	11.8	3.539	0.00011
DE C5 OVHD	130	100	15.2	0.912	0.00011
C4/C5 PROD	145	105	1	4.50E-02	0.00011
STRIPPER REB	206	251	14.1	0.564	0.00021
DE C3 REB	311	339	15	0.962	0.00016
DE C5 REB	353	372	16.6	1.573	0.00016
LCO RCH OIL FRM SPG	125	231	2.3	3.89E-02	9.80E-05
VGO FEED	210	460	8.6	6.19E-02	4.30E-05
VAC RESID FEED	325	460	4.6	6.19E-02	4.10E-05

NOTES
(1) UNITS OF CP ARE IN [MMBTU/HR/F]
(2) UNITS OF ENTHALPY ARE [MMBTU/HR]
(3) UNITS OF HTC ARE [MMBTU/HR/FT^2/F]

Alkylation Unit

The Stratco process, called the " Effluent Refrigeration " process, uses a single-stage reactor in which the temperature is maintained by cooling coils. The reactor contains an impeller that emulsifies the acid-hydrocarbon mixture and recirculates it in the reactor.

Emulsion removed from the reactor is sent to a settler for phase separation. The acid is recirculated and the pressure of the hydrocarbon phase is lowered to flash vaporize a portion of the stream and reduce the liquid temperature to between 20 and 40°F. The cold liquid is used as a coolant in the reactor tube bundle.

The flashed gases are compressed and liquefied, then sent to the depropanizer where LPG grade propane and recycled isobutane are separated. The hydrocarbon liquid from the tube bundle is separated into isobutane, n-butane, and alkylate in the deisobutanizer column. The isobutane is recycled and n-butane and alkylate are product streams[1].

Figure 2.2 is a simple block flow diagram of the process, and Table 2.2 is the stream data used for pinch analysis.

Table 2.2
Atlas Alkylation Unit
Stream Data

STREAM NAME	SUPPLY TEMP. [F]	TARGET TEMP [F]	ENTHALPY CHANGE (2)	MCP (1)	HTC (3)
FDEFIN\FDEF1	105.58	54	6.8	0.132	1.10E-04
2STOUT\CONDOUT	138.4	105	20.7	6.20E-01	1.10E-04
DC3FBX\DC3BTC	160.8	105	3.3	5.91E-02	0.00011
DC3BOT\DC3FBX	202.2	160.8	2.9	7.01E-02	1.10E-04
DEC4BOTP\DEC4BOTC	285.1	105	6.3	3.52E-02	1.10E-04
DC3OHL\C3COOL	117.4	100	0.1	5.75E-03	1.10E-04
DIB OVHD	128.2	119.9	45.4	5.47E+00	1.10E-04
DE C4 OVHD	138.7	133.5	11.5	2.212	0.00011
DE C3 OVHD	120.7	116	16.7	3.55E+00	0.00011
NBUTP\NBUTC	134.4	100	0.6	1.74E-02	0.00011
DIBOHL\DIBOHC	119.9	105	1.6	0.107	0.00011
EFFLP\EFFLW	37.19	95.86	6.8	0.116	0.00011
FVO\DC3IN	104.7	147.8	2.9	6.73E-02	0.00011
TOTWAT\HOTWAT	130.8	139.3	1.4	1.65E-01	0.00018
DIB REBOIL	174.5	207.4	47.3	1.438	0.00016
DE C4 BOTT	245.3	284.7	12.6	0.32	0.00016
DE C3 BOTT	200.9	202.2	20.1	1.55E+01	0.00016

NOTES
(1) UNITS OF CP ARE IN [MMBTU/HR/F]
(2) UNITS OF ENTHALPY ARE [MMBTU/HR]
(3) UNITS OF HTC ARE [MMBTU/HR/FT^2/F]

Utilities

The refinery and utility plant currently produce and consume saturated steam at 425 psig, 125 psig, and 15 psig. A 250 psig header exists, but has no flow in or out. Approximately 210,000 lb/hr of steam is generated at the highest pressure level by 3 boilers (one additional boiler is not kept on line) and a hydrogen plant. The vast

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[1] J. H. Gary and G. E. Handwerk, *Petroleum Refining, Technology and Economics*, 2nd ed. (Marcel Dekkar Inc., N.Y. 1984) pp. 169-172.

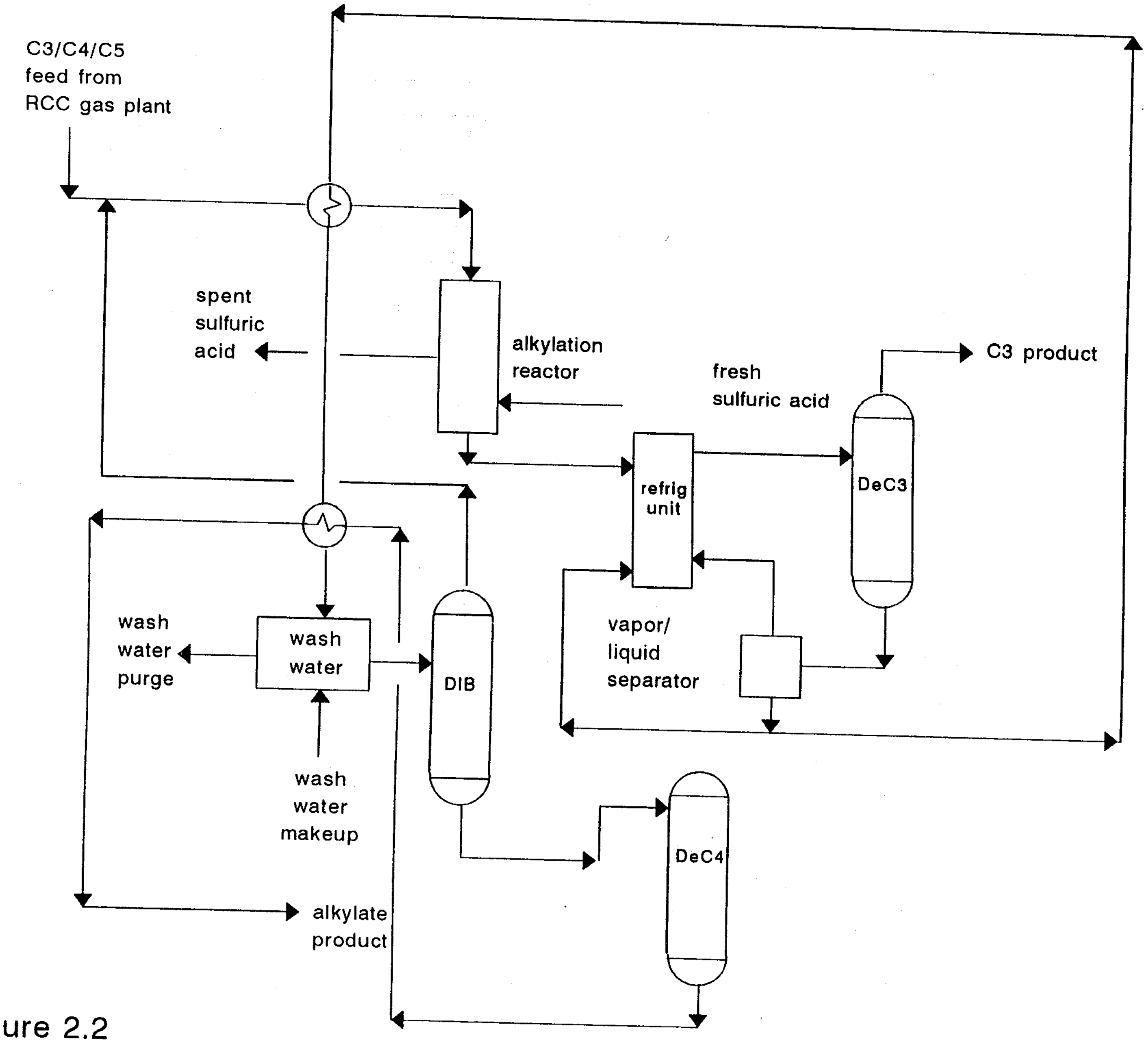

Figure 2.2
Alkylation Plant Flow Scheme

majority of the steam at the other levels comes from let down through valves from the header one pressure level higher. Only 21,000 lb/hr comes from process units and flashing condensate. Information from the site reveals that condensate recovery is low (20% of boiler production) and steam loss from leaks is high. Thus there should be a high return on investment for projects to improve housekeeping. The low recovery of condensate, which is typical of all refineries, requires that utility costs for steam be developed considering both the condensate recovery and no condensate recovery scenarios.

The project design for new units includes generating superheated 425 psig steam at 600°F, as well as the addition of a new 50 psig steam header. There are design options for adding turbines exhausting to either the 125 psig header or the 50 psig header. However, these turbines would have fixed process duties, and the marginal steam flow will continue to be through letdown valves. The marginal costs of utilities are given in Table 2.3. The procedure for calculating these values is discussed in Section 3.3.

Table 2.3
Marginal Utility Costs in Existing Site

	100% Condensate Recovery[a]		**0% Condensate Recovery**[b]	
Steam Pressure	**$/MMBtu**	**$/1000 lb**	**$MMBtu**	**$/1000 lb**
425 # sat	5.48	5.40	8.23	6.34
425 # super	5.45	5.92	7.87	6.84
125 #	5.48	5.34	7.35	6.28
50 #	5.52	5.30	6.84	6.23
15 #	5.53	5.23	6.52	6.16

A steam balance for the existing plant is shown in Figure 2.3. The flows for the individual units are tabulated in Table 2.4.

(a) Assumes recovered condensate is flashed to the 15 psig header. The 15 psig condensate is sent to condensate storage at atmospheric pressure and 190° F. The heat from condensate temperature to storage temperature is lost.

(b) Heat is only recovered down to saturated condensate at the steam pressure.

Data provided by Pennzoil:

Raw water	$1.25/1000 gal
Boiler make up water	$2.60/ 1000 gal
Cooling Water (pumping, makeup water, and chemicals)	$0.20/1000 gal $0. 96/MMbtu
Fuel cost (LHV)	$3.88/MMbtu
Electrical cost	$0.035/KWH

Figure 2.3 Atlas Refinery Steam System: Existing Operation

ATLA_STM.XLS

STEAM SYSTEM MODEL
Case: PENNZOIL
Project: *ATLAS REFINERY (RCC)*

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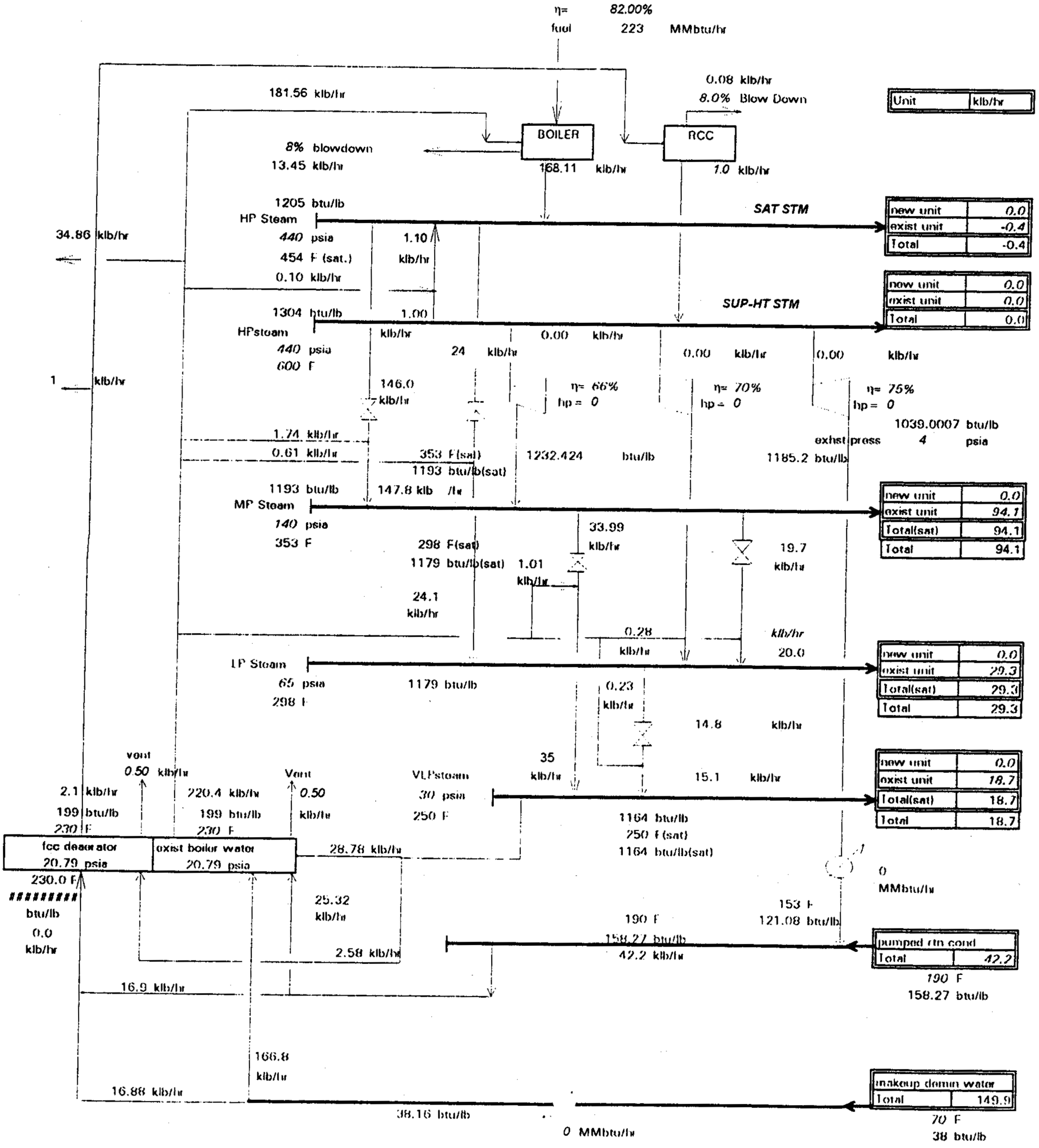


Electrical Power Balance	HP
shaft work	0
Net Import	0
Annual Cost MM $$/yr	0.00

fuel Balance	MMbtu/hr
Boilers	223
Preheat	0
Total	223
	7.12

Water Balance	klb/hr
Demin	150
Total	150
	0.39

On stream time: *8300* hrs/yr

Utility costs

Electricity	*0.035*	$/KWh
Fuel gas	*3.85*	$/MMbtu
Demin wat	*2.6*	$/1000gal

Total energy cost
7.507 MM$/yr

5/18/94

Table 2.4 Atlas Steam Balance with New RCC/Gas Plant Alkylation Unit

	Steam Generation, klb/h						Steam Use, klb/h					
	425#	250#	125#	15#	Total		425#	250#	125#	50#	15#	Total
EXISTING UNITS												
H2 Plant	24.91				24.91							0.00
H2 W.H. Boilers			2.21		2.21							0.00
DDD Steam Gen			7.74		7.74							0.00
Sulfur Recovery Unit			0.00		0.00							0.00
M.E.K.				5.00	5.00		2.10		43.91			46.01
Amine Flash Drum				0.20	0.20							0.00
Condensate Flash Drum				6.00	6.00							0.00
New Flare & Tank Farm					0.00		0.15					0.15
#4 CDU & Stabilizer					0.00		9.03		20.83			29.86
Hydrotreater					0.00		13.31		3.52			16.83
Wax Hydrofinisher					0.00				15.00			15.00
NMP Unit					0.00				0.30			0.30
PDA Unit					0.00				14.83			14.83
Sour Water Stripper					0.00					5.45		5.45
#4 Vacuum Unit					0.00					23.84		23.84
Debutanizer Stripper					0.00				1.19			1.19
Vac Tower Bottoms					0.00				1.00			1.00
Diesel Stripper					0.00				0.00			0.00
40# Amine Unit					0.00				3.44			3.44
PDA Low Temp Evaporator					0.00						9.89	9.89
Losses to Atmosphere					0.00						20.00	20.00
TOTAL	24.91	0.00	9.95	11.20	46.06	0.00	24.59	0.00	104.02	29.29	29.89	187.79
NET							-0.32	0.00	94.07	29.29	18.69	141.73
NEW UNITS												
Alkylation					0.00				10.80	77.00		87.80
FCC/Gas Plant					0.00		18.00					18.00
Sour Water Stripper					0.00					4.35		4.35
Amine					0.00					4.70		4.70
Butamer					0.00		10.00			5.00		15.00
Sulfur Plant			4.00		4.00							0.00
TOTAL	0.00	0.00	4.00	0.00	4.00	0.00	28.00	0.00	10.80	91.05	0.00	129.85
NET							28.00	0.00	6.80	91.05	0.00	-4.00

3

ANALYSIS

Pinch analysis (references 1 and 2) was used to identify energy efficiency improvements for the two new process areas, the RCC/gas plant and alkylation unit. The procedure used was as follows:

i. The optimum value of ΔT_{min} (minimum approach temperature for heat integration) was determined by evaluating the tradeoff between energy and capital costs using the SuperTarget™ program.

ii. "Basic Targets" for improved process heat recovery at the optimum ΔT_{min} were computed.

iii. Options for improving heat recovery by changing the process design and/or operating conditions were explored. The +/- principle of pinch analysis (see reference 2) was used for this part of the analysis.

iv. The interface between the process and the utility system (primarily the steam system) was investigated. Options for preheating Boiler Feed Water (BFW) and Deaerator Feed Water (DFW), and changing steam pressures for specific services, were assessed.

v. Specific instances of cross pinch heat transfer were identified and quantified.

vi. Heat integration matches were defined to correct the cross pinch heat transfer and implement the process modifications and utility changes.

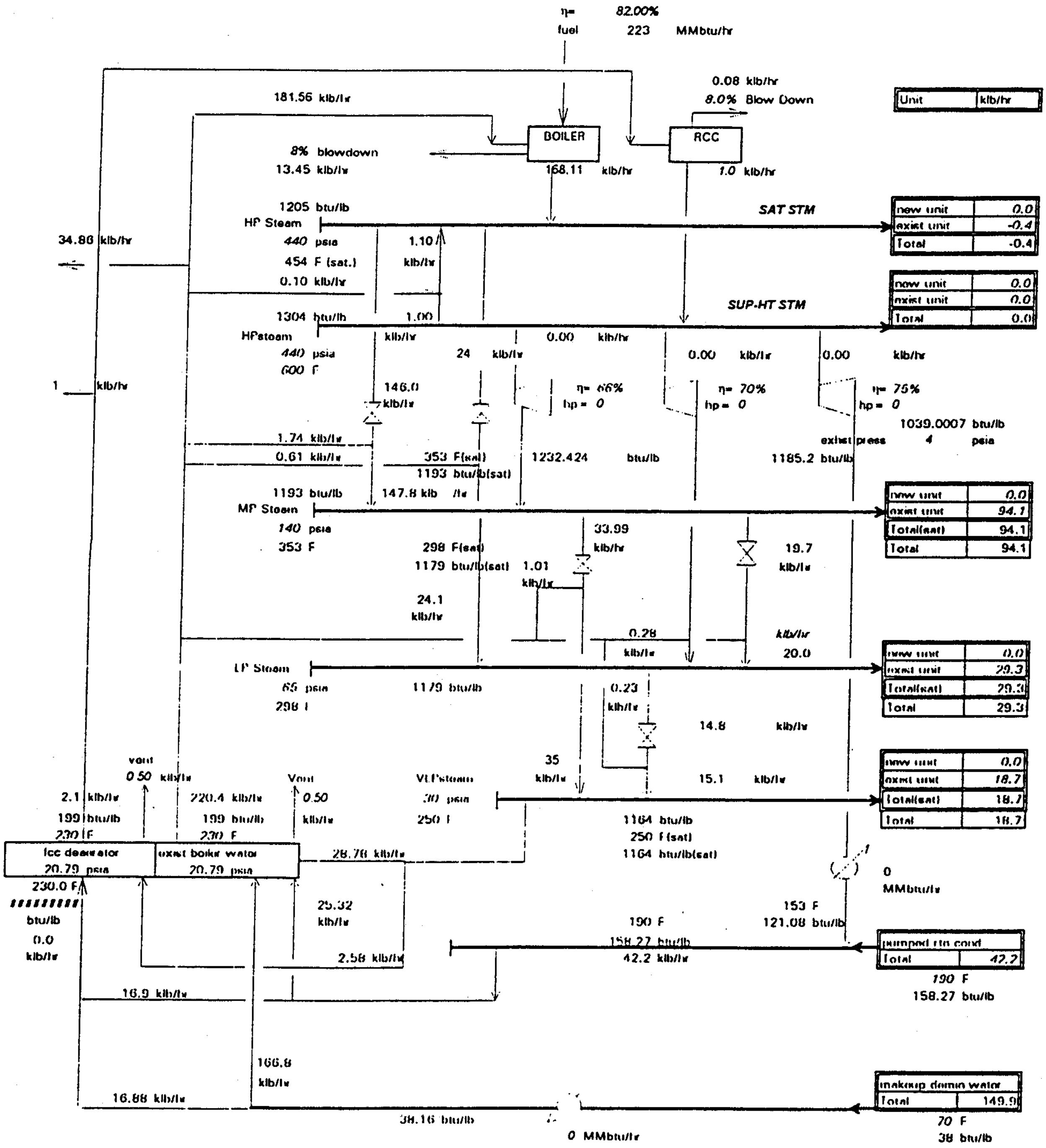
The above procedure was applied to the RCC/gas plant and the alkylation unit. The approach for the steam system was very different. In this case a rigorous spreadsheet model was constructed to simulate the operation of the steam system under various scenarios (see Figure 3.1). The model was coupled to cost data for fuel, electricity, and makeup water, and the annual operating cost was automatically computed for the given scenarios (see bottom right hand corner of Figure 3.1). This spreadsheet provides a mechanism for calculating the cost implications of any change in steam flows and for calculating marginal steam costs. For example, the change in total operating cost when steam flow at a given level changes by 1000 lb/hr is equal to the marginal cost of steam at that level multiplied by the number of operating hours per year (8300).

Figure 3.1 Atlas Refinery Steam System: Existing Operation

ATLA_STM.XLS

STEAM SYSTEM MODEL
Case: PENNZOIL
Project: *ATLAS REFINERY (RCC)*

Units Status OK 0



Electrical Power Balance	HP	fuel Balance	MMbtu/hr	Water Balance	klb/hr
shaft work	0	Boilers	223	Demin	150
		Prohwat	0		
Net import	0	Total	223	Total	150
Annual Cost MM $$/yr	0.00		7.12		0.39

On stream time: *8300* hrs/yr

Utility costs

Electricity	*0.035*	$/KWh
Fuel gas	*3.85*	$/MMbtu
Demin wat	*2.6*	$/1000gal

Total energy cost
7.507 MM$/yr

RCC Unit & Gas Plant

The Cat Cooler and the CO Boiler were excluded from the pinch analysis for this unit. These equipment items produce steam in large quantities and are proprietary designs that were accepted as a constraint of this study.

The results of the targeting analysis considering the rest of the equipment are as follows:

Current steam load =	14.97 MMBtu/h
Basic Target steam load =	10.4 MMBtu/hr
Basic Target Savings =	4.57 MMBtu/h
Value =	$208,000/year
Optimum ΔT_{min} =	20°F

Figures 3.2 and 3.3 show the process composite curves and the grand composite curve, respectively. Table 3.1 is the cross pinch heat exchanger summary table.

The existing design does not achieve the target because of four small pinch violations (see Table 3.1). Consequently, a significant number of changes would be needed to construct an ideal HEN. These changes were considered to be non-viable. In addition, several process modifications were identified, but were found to be unattractive. However, evaluation of the process/utility interface revealed a very attractive option for DFW preheating, summarized below:

Heat Source	Savings (MMBtu/hr)	$'000/yr	Cap Cost $,000	Payback (years)
LCO Product	2.23	96	(69)	n/a
RCC Gasoline	6.71	287	(89)	n/a
Piping (to/from RCC)			225	
TOTAL	8.94	383	67	0.2

The capital cost for this project has been shown as a negative number (except for the piping entry) and payback as "n/a" for both heat sources because the DFW cooler would be a shell and tube heat exchanger with less capital expense than the air cooler it replaces. The capital cost reduction was credited because this is a new design, and the heat exchangers deleted from the base case are not yet in existence. A schematic showing how this project would be integrated into the process is shown as Figure 3.4.

Figure 3.2 Atlas RCC/Gas Plant Composite Curves

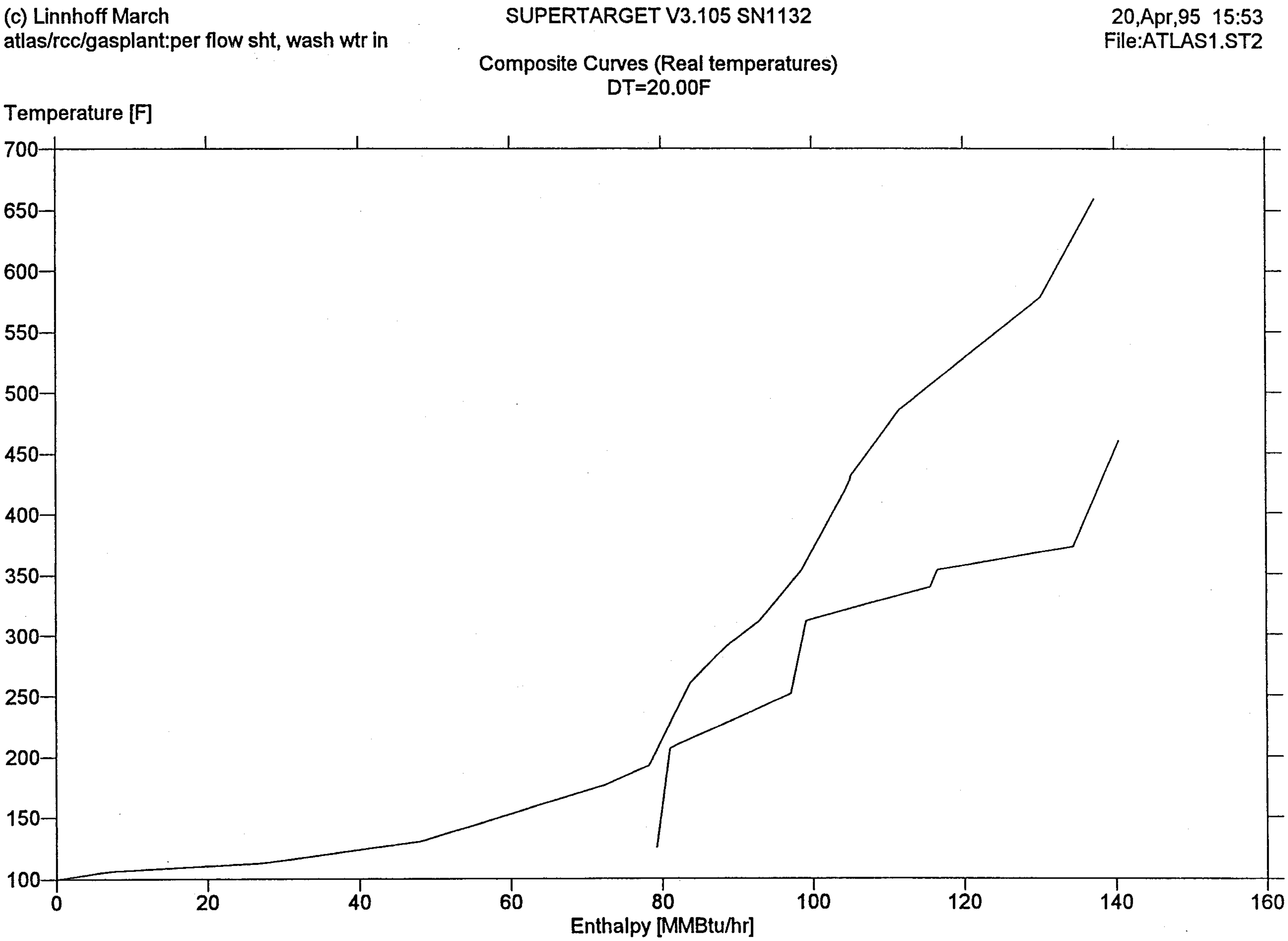


Figure 3.3 Atlas RCC/Gas Plant Grand Composite Curves

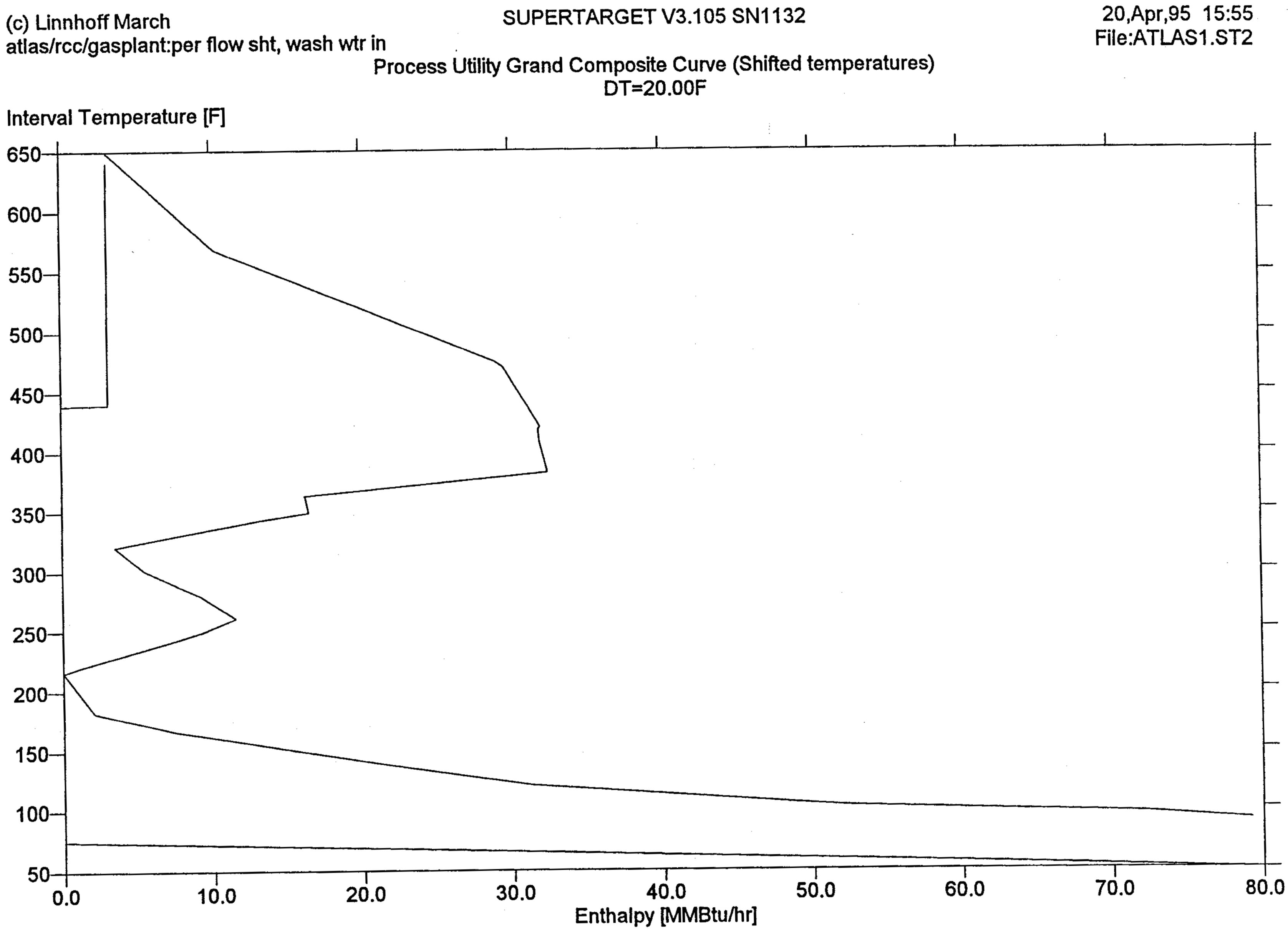


Table 3.1 Atlas RCC/Gas Plant Cross Pinch Table

				ATLAS				DTmin :	20
Atlas RCC/Gas Plant									X-P Ht Tfr
			Hot Side			Cold Side			Process
Equipment	Service Name	Duty	Stream	Ts	Tt	Stream	Ts	Tt	216
Item No.		MMBTU/HR		°F	°F		°F	°F	MMBTU/HR
81-EX-4002	SLURRY P/A STM GEN	5.543	SLURRY P/A	659	485	BFW/425 PSI STM GEN	230	450	0.00
81-EX-1004	RCC CHG/SLURRY P/A EXCH	9.049	SLURRY P/A	659	485	RCC CHG	331.0	460.0	0.00
81-EX-4003	BFW/CLARIFIED OIL COOL	1.727	CLARIFIED OIL	659	280	BFW	230.0	400.0	0.00
81-EX-4004	CLARIFIED OIL EXCH	0.478	CLARIFIED OIL	280	175	CW	85	115	0.25
81-EX-5013	DE C5 REBOILER	16.606	HCO P/A	578	430	DE C5 REBOIL	353.0	372.0	0.00
81-EX-5009	STRIPPER REBOILER	10.088	LCO P/A	428	290	STRIPPER REBIL	219.0	251.0	0.00
81-EX-5011	DE C3 REBOILER	14.970	425# STEAM	454	453	DE C3 REBOILER	311.0	339.0	0.00
81-EX-5006	SPONGE ABS/LEAN OIL EXCH	2.292	LCO TO SPONGE	290	185	RICH OIL	125	231	0.86
81-EX-5007	SPNGE ABS LEAN OIL COOL	1.544	LCO TO SPONGE	185	105	CW	85.0	115.0	0.00
81-EX-1002	RCC CHG/ LCO PROD EXCH	0.952	LCO PROD	417	330	VGO CHG	310.0	338.0	0.00
81-EA-4002	LCO PROD COOL	2.230	LCO PROD	330	125	AIR	85.0	110.0	1.13
81-EX-4005	LCO PROD TRIM COOL	0.270	LCO PROD	125	100	CW	85.0	115.0	0.00
81-EX-5008	STRIPPER REBOILER EXCH	4.016	DE C3 BOTT	311	260	STRIPPER REBOIL	206.0	219.0	0.00
81-EX-1001	RCC CHG/ GASOLINE HTR	3.243	RCC PROD	353	280	VGO CHG	210	310	0.00
81-EA-5003	RCC GASLOINE COOLER	6.710	RCC PROD	280	125	AIR	85	110	2.34
81-EX-5015	RCC GASOLINE TRIM COOLER	1.130	RCC PROD	125	100	CW	85	115	0.00
81-EA-4001	MAIN FRAC CONDENSER	21.670	MAIN FRAC OVHD	192	130	AIR	85	110	0.00
81-EX-4001	MAIN FRAC TRIM COOL	5.040	MAIN FRAC OVHD	130	100	CW	85	115	0.00
81-EX-5001	WGC INTERSTAGE CONDENSER	3.534	WGC INTERSTAGE	130	100	CW	85.0	115.0	0.00
81-EA-5001	HP RECEIVER CONDENSER	6.144	WGC DISCHRGE	176	130	AIR	85	110	0.00
81-EX-5003	HP RECEIVER TRIM CONDENSER	5.796	WGC DISCHRGE	130	100	CW	85.0	115.0	0.00
81-EX-5004	PRIMARY ABS TOP INTERCOOLER	0.726	ABS TOP P/A	118	100	CW	85.0	115.0	0.00
81-EX-5005	PRIMARY ABS BOTT INTERCOOLER	0.525	ABS BOTT P/A	112	106	CW	85	115	0.00
81-EX-5010	DE C3 COND	11.796	DE C3 OVHD	112	106	CW	85.0	115.0	0.00
81-EX-5012	DE C5 CONDENSER	15.207	DE C5 OVHD	130	100	CW	85.0	115.0	0.00
81-EX-5014	C4/C5 PROD COOL	1.000	C4/C5 PROD	145	105	CW	85.0	115.0	0.00
				RCC/Gas Plant cross-pinch heat transfer total					4.57

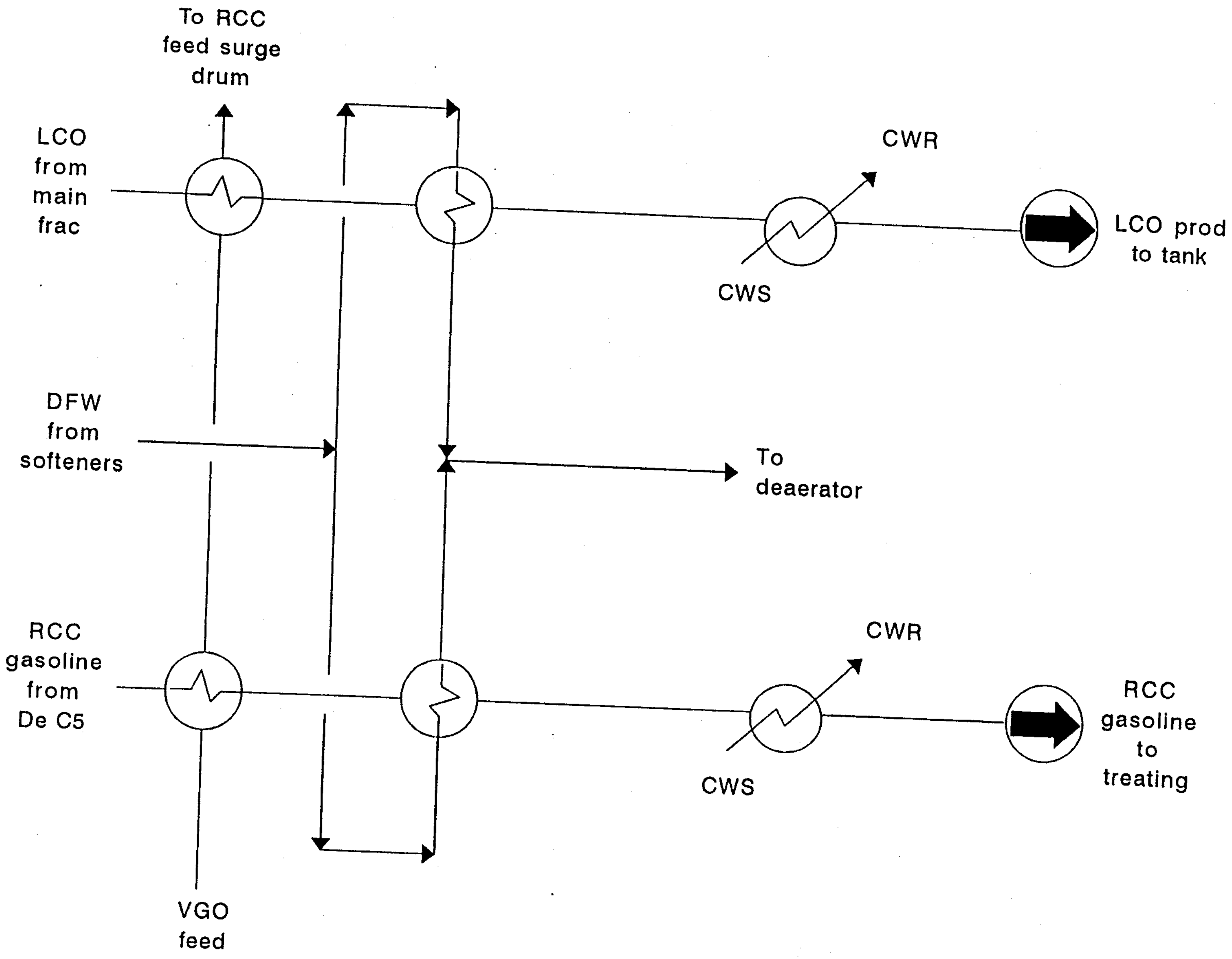


Figure 3.4
DFW Preheating Options

Alkylation Unit

The results of the basic pinch targeting for the alkylation unit are:

Current steam load =	79.7 MMBtu/h
Basic Target steam load =	76.3 MMBtu/h
Basic Target savings =	3.4 MMBtu/h
Value =	\$156,000/year (50 # stm, no condensate return)
Optimum ΔT_{min} =	10°F

In addition to the summary presented above, Figures 3.5 and 3.6 show the process composite curve and the grand composite curve, respectively. Table 3.2 shows the cross pinch heat exchanger summary table.

The preliminary design did not achieve the target because of three small pinch violations (see Table 3.2). The changes required to meet the target were judged to be economically unattractive.

The following process modification options were identified:

1. Add another shell to the DeC3 feed/bottoms exchanger. The preliminary PFD shows that there is an LMTD (uncorrected) of 55.3°F for the feed/bottoms exchanger. The feed to the DeC3 column is highly subcooled with this design. By providing sensible heat to the feed, duty in the reboiler could therefore be reduced. Simply by adjusting the approach of this exchanger to the optimum ΔT_{min} of 10°F, an additional 2.8 MMBtu/hr can be provided to the feed. Simulations indicate that this would reduce the reboiler duty by 2.8 MMBtu/hr.

2. Use of the DeC4 bottoms heat. The pinch analysis highlighted the inefficiency arising from the cooling of the DeC4 bottoms against wash water (a cross pinch duty). The energy consumption could be reduced if the cooling was to be done in such a manner as to decrease the overall utility demand. Both of the following projects accomplish this objective. However, it is not possible to do both projects. They are mutually exclusive. Projects which are mutually exclusive are identified in this report by the use of a common identifying number, but different identifying letters (e.g. 2a,2b, 2c, etc.).

   a. Heat the main DIB column feed. Since the main column feed is subcooled as it enters the DIB column, it is possible to reduce the reboiler load by bringing the feed up to its bubble point. Computer simulations of the DIB column show that this general rule applies in this case and there is a 1:1 relationship between feed preheating and reboiler savings. The

Figure 3.5 Atlas Alkylation Unit Composite Curves

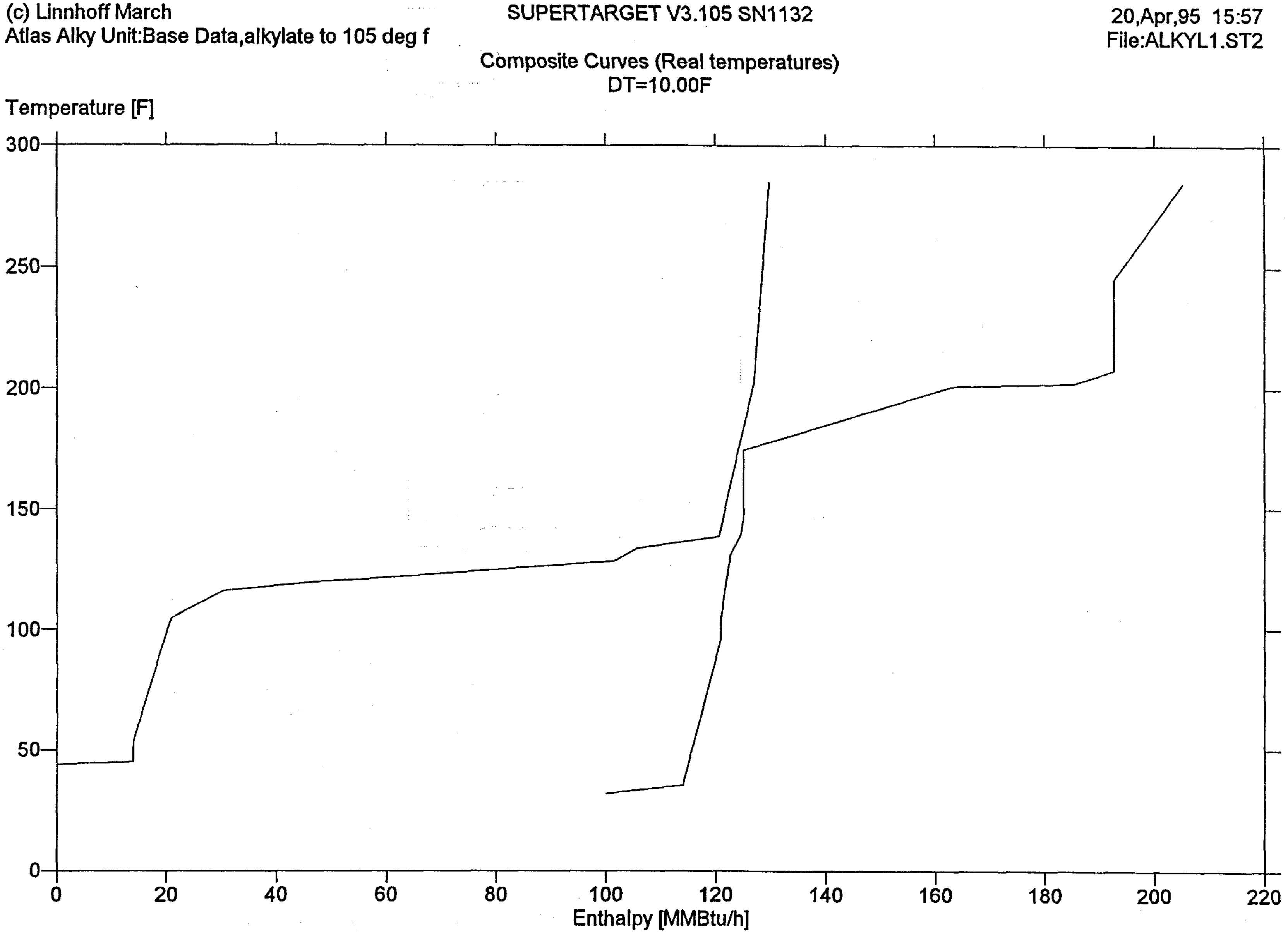


Figure 3.6 Atlas Alkylation Unit Grand Composite Curves

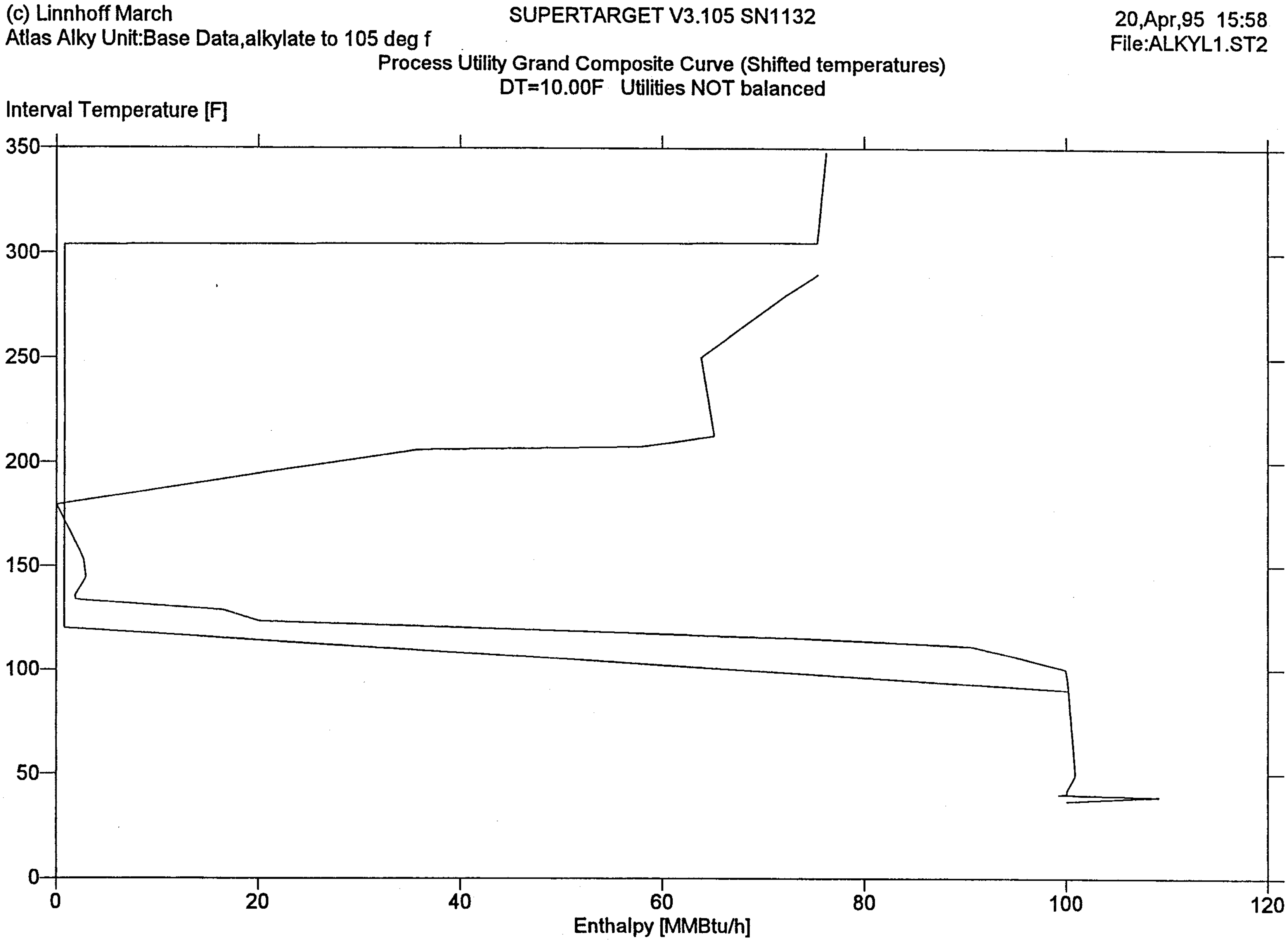


Table 3.2 Atlas Alkylation Unit Pinch Table

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ATLAS ALKYLATION UNIT							DTmin :	10
								X-P Ht Tfr
		Hot Side			Cold Side			Process
Equipment	Duty	Stream	Ts	Tt	Stream	Ts	Tt	189.5
Item No.	MMBtu/h		°F	°F		°F	°F	MMBtu/h
1	6.805	FDEFIN\FDEF1	105.58	54	EFFLP\EFFLW	37.19	95.86	0.00
2	14.053	NBUNQ	45	44	BUNIN\BUNOUT	31.8	35.2	0.00
3	20.660	2STOUT\CONDOUT	138.4	105	REFCON-CWS\CWR	85.0	115.0	0.00
4	3.345	DC3FBX\DC3BTC	160.8	105	DC3BCL-CWS\CWR	85	115	0.00
5	2.910	DC3BOT\DC3FBX	202.4	160.8	FVO\DC3IN	104.7	147.8	0.55
6	1.150	DEC4BOTP\DEC4BOTC	273.4	245.3	TOTWAT\HOTWAT	130.8	139.3	1.15
16	4.769	DEC4BOTP\DEC4BOTC	245.3	105.0	CWS\CWR	85.0	115.0	1.73
7	0.092	DC3OHL\C3COOL	116	100	C3PCL-CWS\CWR	85	115	0.00
8	44.900	DIB OVHD	132.4	124.1	CWS\CWR	85.0	115.0	0.00
9	47.800	50 # STM	298	297	DIB REBOIL	184.9	226.0	0.00
10	9.150	DE C4 OVHD	130.5	125.1	CWS\CWR	85.0	115.0	0.00
11	9.366	125# STM	353	352	DE C4 BOTT	237.6	273.4	0.00
12	16.700	DE C3 OVHD	120.7	116	CWS\CWR	85.0	115.0	0.00
13	20.200	50 # STM	298	297	DE C3 BOTT	201.1	202.4	0.00
14	2.270	50 # STM	298	297	NEW ADDITION	201.1	202.4	0.00
15	0.438	NBUTP\NBUTC	125.1	100	NC4CL-CWS\CWR	85	115	0.00
16	2.050	DIBOHL\DIBOHC	124.1	105	DIBOHCL-CWS\CWR	85	115	0.00
			Alkylation Plant cross-pinch heat transfer total					*3.43*

feed is below the pinch temperature. Feed preheating can therefore be accomplished without increasing hot utility requirements The quantity of heat which can be added as feed preheat is limited by the need to maintain the required ΔT_{min} on the downstream wash water preheat exchanger (see Figure 3.7). This change will save 2.88 MMBtu/hr, worth $132,000/yr. Its incremental capital cost is only $45,000, which translates to a payback of 0.34 years.

b. Heat sodium zeolite softened city water for wash water makeup. 15,000 lb/hr of wash water make up is required in the alkylation process. In the base case design this make up consists of hot condensate. However, it would be possible to substitute cold sodium softened water and use the heat from the DeC4 bottoms to compensate for the lower temperature of this stream compared to condensate. There is enough heat in the DeC4 bottoms to allow sodium softened city water to be used. The only optimization necessary is that the ΔT_{min} of the exchanger be reduced to 10 °F for the winter operation mode. This change achieves roughly the same energy savings as project 2a (2.73 MMBtu/hr versus 2.88 MMBtu/hr). However, its savings are greater ($137,000/yr) because it also saves treating cost for condensate . The incremental capital cost and payback for this project are $29,000 and 0.21 years, respectively.

3. Inter-reboil the DIB column by process heat integration. Two mutually exclusive projects based on this concept are:

a. Use an MVR heat pump on the DeC4 overheads and utilize the DIB side reboiler as the DeC4 condenser. This project utilizes a relatively small (285 hp) compressor to boost the pressure of the DeC4 overheads so that they can condense against a side reboiler of the DIB column. Figure 3.6, the grand composite curve for this unit, reveals that this heat pump elevates heat across the process pinch. In pinch technology terms, use of a heat pump across the pinch is very desirable. The result is a decrease in both hot and cold utilities.

This project has two advantages over alternative methods. The first advantage is that the design of the DeC4 column need not change significantly, if at all. Both columns remain at the existing design pressures. Therefore all current simulations are still valid. The other advantage is that there is an interest on the part of the Department of Energy (DOE) to encourage heat integration using this principle. Therefore, funding should be available from the DOE to offset some of the capital cost of this project. A schematic of this project appears as Figure 3.8. The energy and monetary savings, capital cost and payback are 11 MMBtu/hr, $436,000/yr, $1.4 MM, and 3.2 years, respectively. Note that

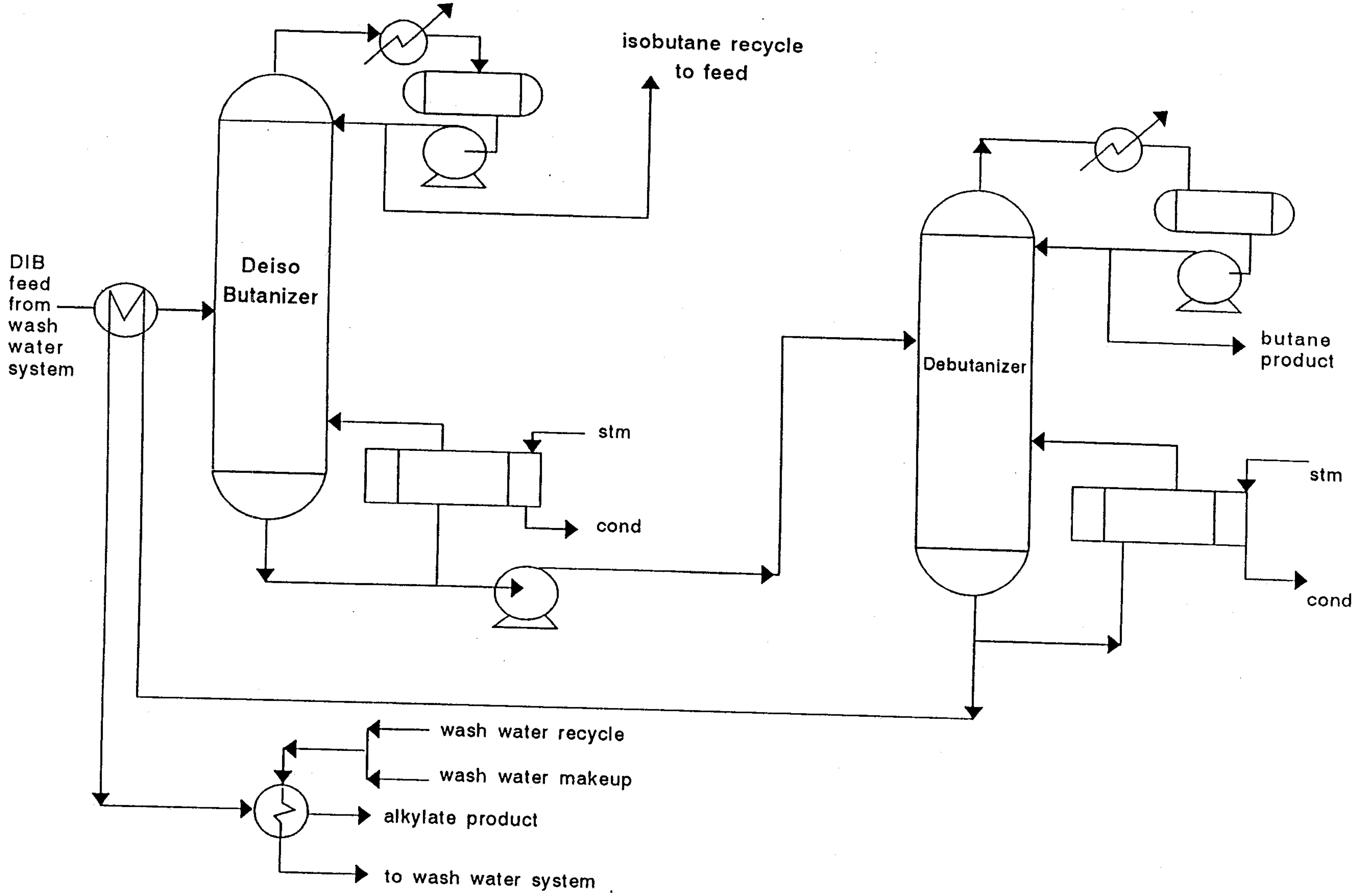


Figure 3.7
Schematic of Debutanizer Bottoms
forDIB Column Feed Preheat

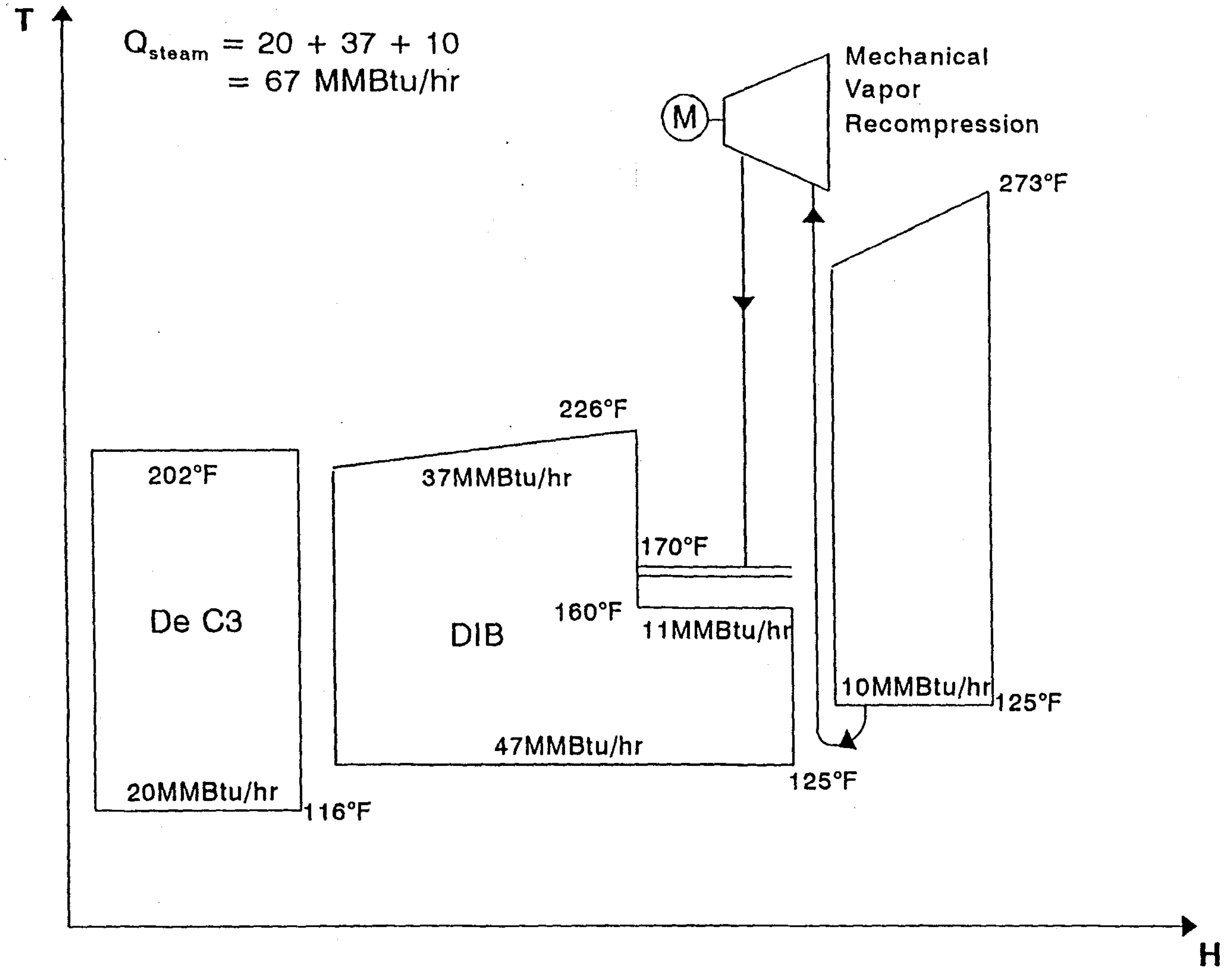


Figure 3.8
MVR Option Temperature
Enthalpy Representation

Preliminary Evaluation:

Power Consumption = 235 kW or \$69,000/yr
Steam Saving = 11 MMBtu/hr or \$504,000/yr
Net Saving = 504 - 69= k\$436/yr
Equipment Cost ≈ MM\$1.4 Payback = 3.2 years

these are very crude estimates, and will require revision if this project is pursued.

b. Raise the pressure of the DeC4 column so the overheads can condense against the DIB side reboiler. Currently the DeC4 column is designed to operate at 95 psia. However, if the pressure is raised to 170 psia, the overheads could condense at 180°F. This would enable heat from the condensing overheads to be used for a DIB side reboiler. Because the DeC4 column is downstream of the DIB column and the side reboiler is only 25% of the total column reboiler duty, startup of the column should not be difficult. However, some safety factors should be provided in the design of the columns to dampen out any swings in one column from affecting the other column's operation. Without the safety factors, the energy and monetary savings, incremental capital cost, and payout are 8 MMBtu/hr, $367,000/yr, $192,000, and 0.52 years respectively. Figures 3.9 and 3.10 show what the project does to the fractionation process on temperature versus enthalpy diagrams. Figure 3.11 show a process schematic along with some feed preheat projects.

The elevation in column pressure elevates the reboiler temperature so that it can no longer utilize 125 psig steam. 425 psig steam might be used instead. However, this should not change the operating cost of the column provided the condensate is recovered (see Table 2.3).

Steam System

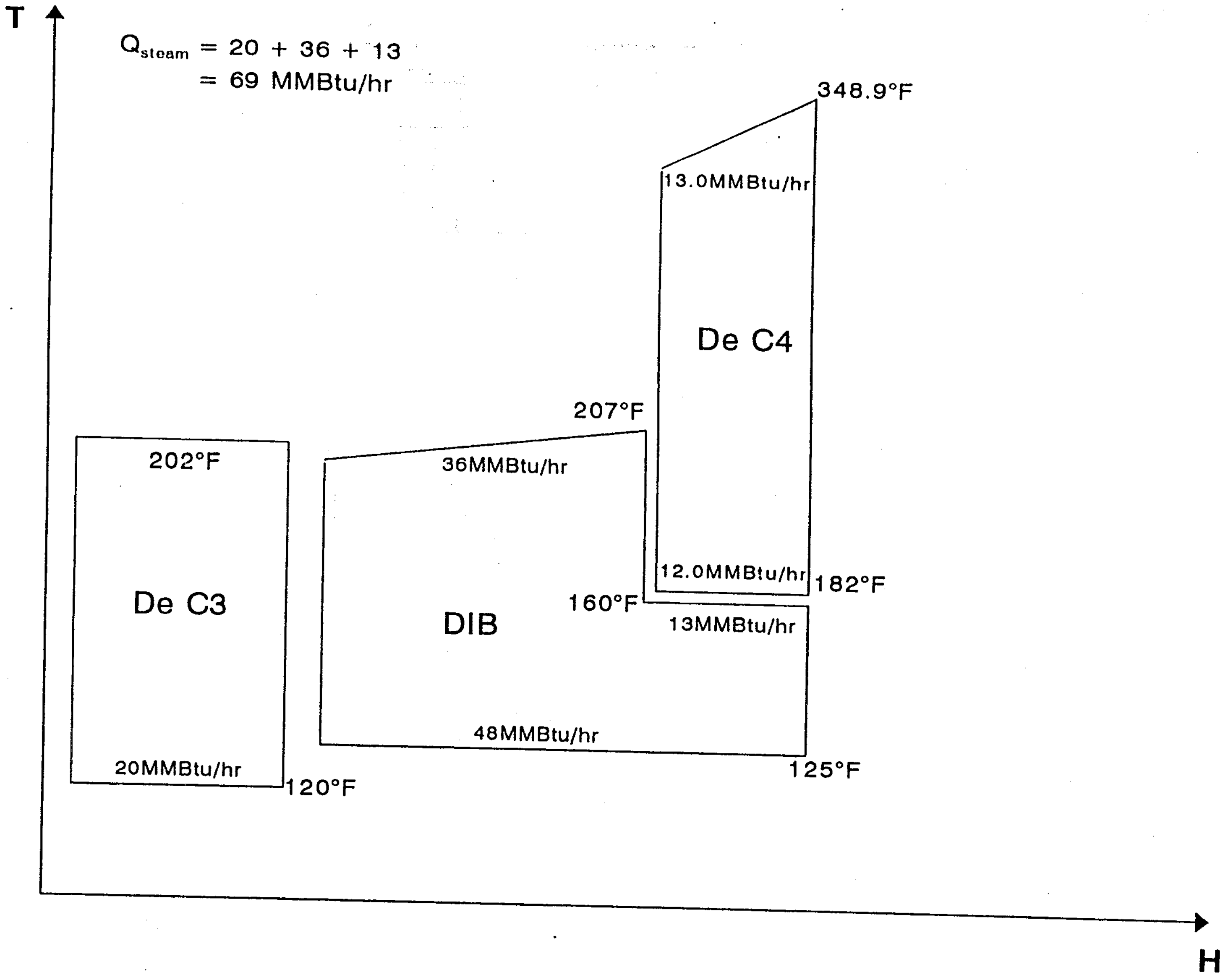
The steam system is the area with the greatest opportunity for optimization. Since the steam system is integral with process units, many of the issues regarding the steam system have already been mentioned. The main opportunities and issues addressed by the study are:

- Selection of steam turbine versus motor drivers for rotating equipment
- Improvement of housekeeping and its impact on payback of various projects

Selection of Steam Turbine Drivers Versus Motor Drivers for Rotating Equipment

In general there are two main reasons that steam turbines are often incorporated in process plants:

1. Reliability-especially where electric supply is known to be suspect.
2. Reducing operating costs. The power from backpressure turbines is generally cheaper than imported electricity.



$$Q_{steam} = 20 + 36 + 13$$
$$= 69 \text{ MMBtu/hr}$$

Figure 3.9
High Pressure DeC4 Column
Temperature Enthalpy Representation

<u>Preliminary Evaluation</u>:
Net Steam Saving = 77 - 69 = 8 MMBtu/hr or <u>k$367/yr</u>

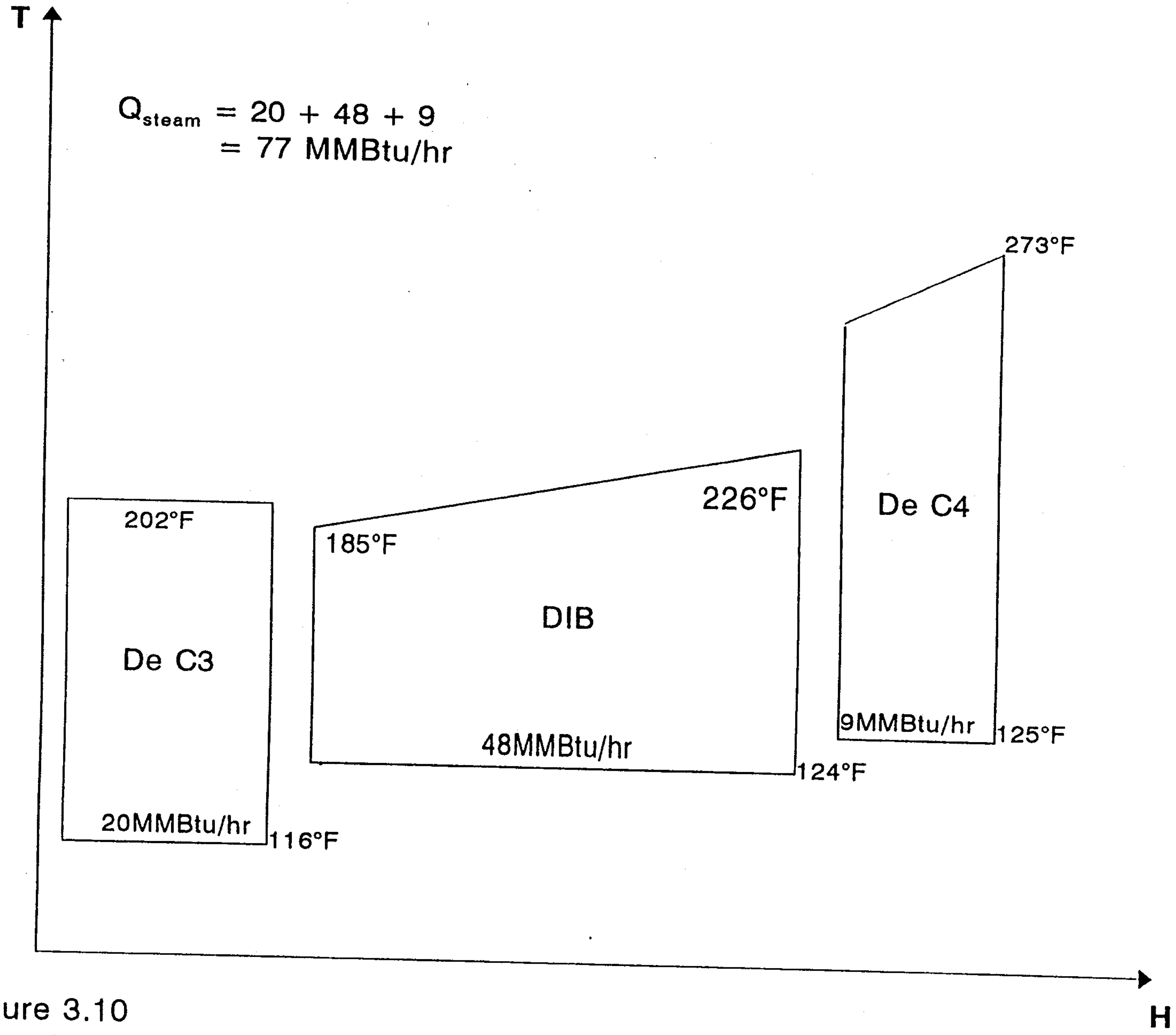


Figure 3.10
Base Case Fractionation Train Design
Temperature Enthalpy Representation

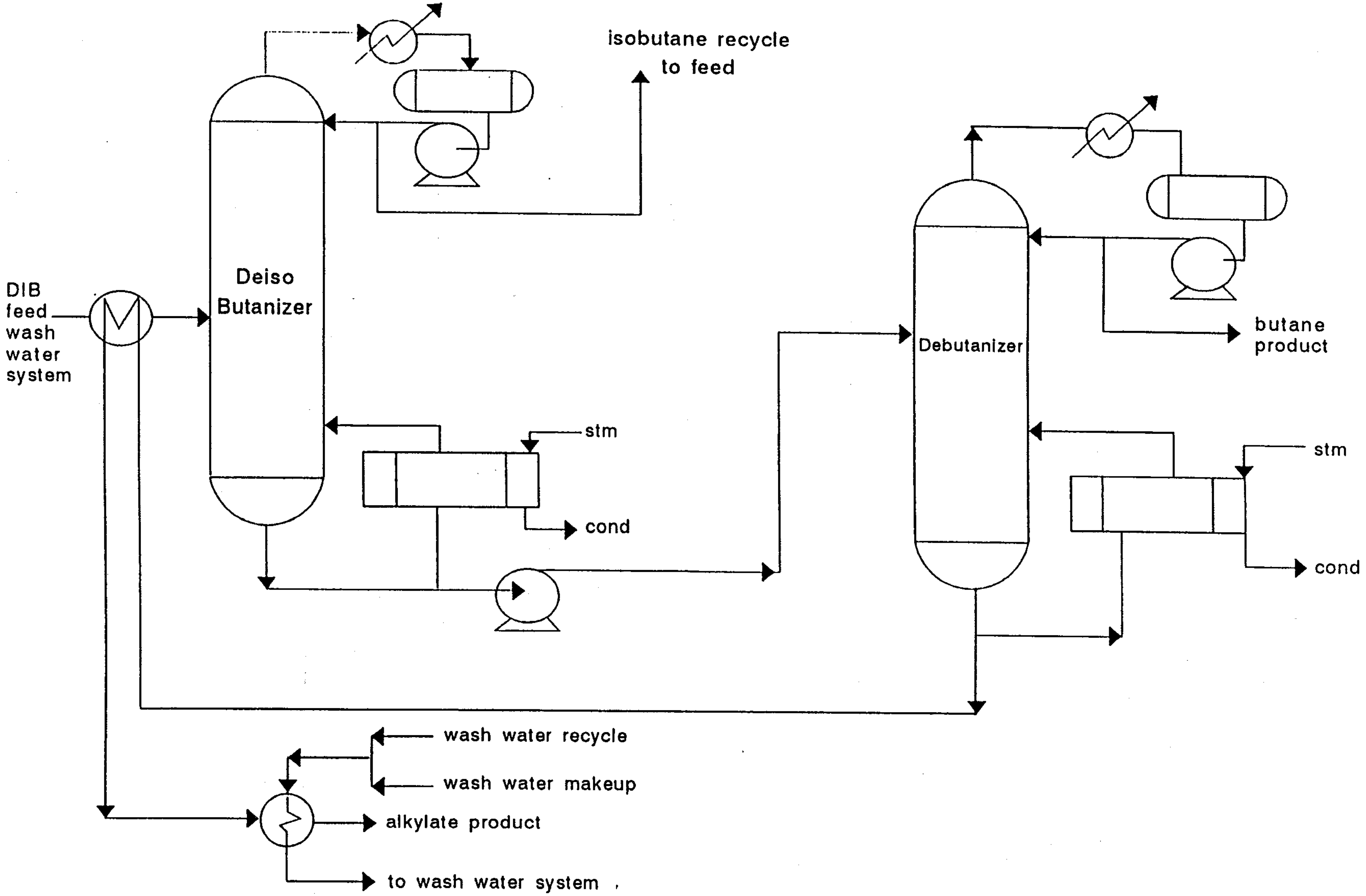


Figure 3.11
Schematic of Debutanizer Bottoms for DIB Column Feed Preheat

Reliability of the SWEPCO electric supply is not a problem at the Atlas site. The only reason for using steam turbines, therefore, is if backpressure turbines would reduce operating costs. In order for this to occur, the following two criteria must be met:

- There needs to be sufficient high pressure steam of suitable quality for the inlet side of the turbine.
- There must be sufficient demand for the steam at the exhaust pressure level.

As part of their design study for refinery expansion, Pennzoil considered installing up to five steam turbines:

- A 4400hp driver for the RCC air blower
- A 2300hp driver for the Alkylation unit's refrigeration compressor
- A 2300hp driver for the Wet Gas compressor in the RCC unit
- Two 500hp drivers for the mixers in the Alkylation unit reactors

At the time the study was carried out various options for backpressure and condensing turbines were under consideration. Linnhoff March re-evaluated these in the light of the pinch analysis and the steam model. The results are summarized in Table 3.3.

The selection of turbines at the Atlas site is limited by both of these considerations. On the availability of high pressure steam, the only suitable (superheated) steam is from the new Cat Cooler and CO Boiler. This amounts to 124,000lb/hr. From Table 3.3, it is clear that this is insufficient for a 4400 hp turbine (to drive the main air blower) if the machine exhausts to 125 psig. The required steam flow is 147,400lb/hr for such a design. However, if the turbine exhausts to 50 psig, the water rate is more favorable, and the steam demand for the turbine is less than the available supply (94,160lb/hr versus 124,000lb/hr). It is not possible to run a 4400 hp and a 2300 hp turbine simultaneously, however, even if they both exhaust at 50 psig, because their combined demand (144,760lb/hr) exceeds the available supply (124,000lb/hr). For operational reasons, Pennzoil favors the selection of a 4400 hp turbine for the air blower, and this necessitates the installation of a machine exhausting at 50 psig, and the introduction of a 50 psig header as part of the expansion.

There are no users of 50 psig steam in the existing Atlas site. However, the DIB and DeC3 columns in the new alkylation unit would use steam at this pressure level. Together they consume 77,200 lb/hr in the existing design, but this will be reduced to 65,300 lb/hr with the changes described in Section 3.2.

This new demand is considerably less than the steam flow through the 4400 hp turbine exhausting to the 50 psig header (94,160 lb/hr). The excess 50 psig steam could be let down to the 15 psig header, but this arrangement is thermodynamically undesirable. Moreover, the net demand for 15 psig steam is small, so the amount of steam that could be let down is limited. It was therefore necessary to identify some existing users

Table 3.3 Steam Turbine Economics

CASE #	HP	DISCHARGE PRESSURE	STEAM FLOW	PURCHASED COST OF TURBINE	INSTALLED COST OF TURBINE	INSTALLED COST OF MOTOR	INCREMENTAL CAPITAL COST	ANNUAL FUEL COST	ANNUAL POWER SAVINGS	NET ANNUAL UTILITY SAVINGS	PAYOUT
1	4400	140	147400	950000	1425000	653600	771400	456500	967800	511300	1.508703
2	4400	60	94160	950000	1425000	653600	771400	493000	967800	474800	1.624684
3	2300	140	81650	825000	1237500	408500	829000	238600	505900	267300	3.101384
4	2300	60	50600	775000	1162500	408500	754000	259600	505900	246300	3.061307
5	500	140	25500	55000	104500	103500	1000	51900	110000	58100	0.017212
6	500	60	16000	50000	95000	103500	-8500	56500	110000	53500	-0.15888
7	4400	8.5 IN HG	42240	850000	1275000	653000	622000	2151000	967800	-1183200	-0.52569
8	2300	8.5 IN HG	23460	800000	1200000	408500	791500	1124400	505900	-618500	-1.27971

Notes

1. Turbines discharging at 140 psig exhaust to the 125 psig header
2. Turbines discharging at 60 psig exhaust to the 50 psig header

of 125 psig steam that could be converted to 50 psig steam use. The existing sour water stripper and #4 vacuum unit, for example, could be converted and would provide the necessary sink for 50 psig steam.

The economics of utilizing steam turbines to produce cheaper shaft work than imported electric are shown in Table 3.3. In most cases, the installed motor cost is less than the installed turbine cost. Thus the credit for cheaper shaft work must offset the cost for increased capital. The condensing turbine does not produce cheaper shaft work than imported electricity. So much energy is discarded in the surface condenser that the fuel costs to revaporize the condensate makes this arrangement uneconomic. One instance where such a turbine might be viable is when the supply of steam is very limited and the integrity of the electrical supply is suspect. The economics of the condensing turbine are even worse than presented in the table because the capital costs do not include the surface condenser and related equipment.

Improvement of Housekeeping and its Impact on Payback of Various Projects

Opportunities were identified to improve the efficiency of Pennzoil's Atlas site with little investment by improving the routine maintenance of the site steam system. Areas for improvement include:

- Reducing steam leaks
- Recovering condensate where economics show a payback
- Investigating the exact cause of the high flow of steam into the deaerator and eliminating any excess steam venting associated with this mode of operation
- Reinstating the boiler blowdown/DFW preheat exchanger.
- Eliminating the frequent venting of 15 psig steam. This is believed to be the result of a control problem in the steam system.

References

1. *Pinch Technology: A Primer.* Electric Power Research Institute, Palo Alto, CA: March 1990. Final Report, EPRI CU-6775.
2. B. Linnhoff, "Pinch Analysis-A State-of-the-Art Review", *Trans IChemE*, Part A, Vol. 71, pp. 503-522, September 1993.

4

CONCLUSION

Results

The pinch analysis resulted in the recommendation of the following projects:

Description	Savings $'000/yr	Capital Cost $'000	Payback yr
1. Blowdown Exchanger	150	n/a	n/a
2. 50 # Steam Turbine (Air Blower)	475	771	1.62
3. RCC +LCO/DFW Preheat	383	67	0.2
4. DeC3- Increased Feed Heat	128	131	1.02
5. Uses for DeC4 Bottoms **(either a or b)**			
5a. City Water for Wash*	137	29	0.21
5b. DIB Feed Preheat	132	45	0.34
6. Intermediate Reboiler on DIB Column **(either a or b)**			
6a. Raise DeC4 Pressure*	367	192	0.52
6b. MVR on DeC4 Overheads	436	1,400	3.2
Total including * options	1640	1,190	0.73

A comparison of the existing design, theoretical targets, and operation of the new units incorporating the recommended projects appears below:

RCC/Gas Plant

Current steam load = 14.97 MMBtu/h
Basic Target steam load = 10.4 MMBtu/h
Basic Target savings = 4.57 MMBtu/h
Value = $208,000/year
Net steam load after projects = 6.03
Value = $383,000/year

Alkylation Unit

Current steam load =	79.7 MMBtu/h
Basic Target steam load =	76.2 MMBtu/h
Basic Target savings =	3.4 MMBtu/h
Value =	$156,000/year
Net steam load after projects =	66.1 MMBtu/h
Value =	$632,000/year

Steam System

Annual Operating Cost, new units, no turbines	$6.93 MM/yr
Annual Operating Cost, new units, with turbine, with housekeeping	$5.63 MM/yr
Annual Operating Cost, new units, with all pinch recommendations above	$4.70 MM/yr

As part of their study, Pennzoil considered installing up to five steam turbines. Specifically under consideration were:

- A 4400hp driver for the RCC air blower
- A 2300hp driver for the Alkylation unit's refrigeration compressor
- A 2300hp driver for the Wet Gas compressor in the RCC unit
- Two 500hp drivers for the mixers in the Alkylation unit reactors

If Pennzoil had installed all five steam drivers, 7457 kW of electric power would have been replaced by steam. However, the pinch analysis demonstrated that only the 4400hp steam turbine driver is economic. This replaces approximately 3300kW of electricity.

The savings exceed the basic targets in the above summary. This is because the pinch projects come from either process/utility system integration or from process modifications. The basic targets only identify what is possible from the original process enthalpy and temperature data. Process modifications and utility integration change the stream data using the +/- principle and reduce the inherent energy requirement of the process.

Recommendations

It is strongly recommended that all six projects listed above should be implemented. For the projects where there are mutually exclusive choices the following options are preferred:

- •5a Use of city water for alkylate wash makeup. The advantages are that the savings are greater, capital cost is lower, and the impact of the DeC4 column operation on the DIB column operation will be reduced.
- •6a Raising the pressure of the DeC4 column to inter-reboil the DIB column. The advantages are reduced capital cost and less dependence on rotating equipment. However, the potential availability of DOE funding to offset some of the capital cost could make the MVR on the DeC4 overheads (option 6b) more attractive and bring it back into consideration.

At the time of writing all of the above options are being reviewed by Pennzoil.