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energy conservation in distillation

A Technology Applications Manual
Technology Transfer

U.S. Department of Energy

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Technology Transfer

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Chapter 1 Introduction

This technology applications manual is intended to assist process engineers in technical and economic analyses of the energy-conservation measures most suitable for their particular distillation applications. Its specific objectives are to:

- Catalog all the energy-conservation options applicable to distillation
- Catalog the options by the investment required
- Describe in detail the options having a significant potential to reduce distillation energy requirements economically
- Provide guidelines that will allow the plant engineer to quickly screen each option for his application
- Provide short-cut calculation procedures for use in a preliminary economic analysis of promising options

A companion to this report, the executive briefing report *Energy Conservation: A Route to Improved Distillation Profitability*, has been prepared to call the attention of industrial managers and executives to potential energy savings.

Distillation is the most widely practiced technique for separating mixtures of chemical species. Its applications range from the separation of heavy petroleum fractions to the separation of liquefied air. Our society depends on distillation to provide the abundant energy and synthetic materials so essential to our modern life-style.

By its very nature, distillation is an energy intensive process. Components with different boiling points are separated by a series of fractional vaporization and condensation steps. The boiling of liquid components, with the associated high latent heat of vaporization, is an indispensable part of distillation. But, although distillation will always require an energy input, the efficiency with which the energy is used can be greatly improved.

POTENTIAL FOR ENERGY CONSERVATION

It is very difficult to account for all the energy used in distillation because of its broad and diverse applications. The bulk of distillation energy is consumed in the petroleum, chemical, and natural gas processing industries, however. A recent study based on these industries has estimated the annual national distillation energy consumption to be about 2 quads (a quad is 10^{15} Btu). This represents 3% of the total U. S. energy consumption, or the equivalent of 315 million barrels of fuel oil per year.¹

A 10% reduction in distillation energy consumption would effect a significant savings. On a national basis this

would be an annual savings of 200 trillion Btu, or the equivalent of 36.5 million barrels of oil per year. If energy is valued at \$2.50/ 10^6 Btu, the economic benefit would be \$500 million annually.

The technology to achieve these savings in distillation energy is available and has been for many years. Proper distillation design is a compromise between energy consumption and the capital cost of equipment necessary to recover that energy. Energy-conservation programs must compete with other programs for the available capital, and they must show an attractive return on investment to compete successfully.

Historically, energy prices have been so low that only rudimentary conservation efforts could be justified. In recent years that situation has changed rapidly, as can be seen in Fig. 1. The price of energy has increased far faster than the cost of equipment. This implies that most existing distillation facilities, even those built as recently as 2 years ago, can justify energy-conservation expenditures.

SOURCES OF HEAT LOSS

Heat energy is applied to the distillation column through the feed and/or the reboiler. This heat can be provided by heat exchange with other process streams, steam, or circulating hot oil or by a fired heater.

Heat energy leaves the distillation system in the product streams and the overhead condenser cooling medium. Heat is also lost by radiation and convection to the atmosphere and as a result of inefficiency in the heat sources (such as steam condensate or flue gas from a fired heater). One of the oldest and most effective energy-conservation schemes associated with distillation is the recovery of heat from the products by heat exchange with the feed.

The reflux is shown as a heat loss in Fig. 2. The reflux stream recycles inside the system, removing heat by vaporization in the tower and transferring heat to the overhead coolant in the condenser. It is one of the primary factors controlling the overall heat balance in the tower and can be a major source of inefficiency.

DESIGN RATIONALE

The reboiler and the reflux provide the driving force for distillation. The reboiler heats the bottom of the tower while the reflux cools the top. Vapor from the bottom rises until cool enough to condense. Liquid falls until hot enough to vaporize. The result is a column temperature profile such as that shown in Fig. 3.

The concentration of components tends to follow this profile, with the lighter (low-boiling) compounds becoming more concentrated in the upper (cooler) sections of the tower, and the heavier (high-boiling) components concentrating in the lower (hotter) section. The extent to which the concentration profile follows the temperature profile is determined by the efficiency of contact between the vapor and liquid phases. Devices such as trays or packing material are used to promote contact by increasing the area of vapor-liquid interface.

Distillation can be considered as a series of equilibrium flashes, each taking place at a different temperature. This concept is illustrated in Fig. 4. In this case each flash vessel represents a true equilibrium stage. A theoretical plate is a single contacting device that acts as an equilibrium stage, so that the vapor and the liquid leaving the plate are in equilibrium.

As the number of theoretical plates (equilibrium stages) in a distillation column increases, the reflux (and consequently the energy) required to make a given separation decreases. The bounding limits are the minimum number of theoretical plates at total reflux and the minimum reflux with an infinite number of theoretical plates. This relationship is shown graphically in Fig. 5. The optimum reflux

Fig. 2 Factors resulting in heat loss and inefficiency in a simple distillation column.

Fig. 1 Escalation of fuel and equipment costs.

Fig. 3 Temperature profile of a distillation column.

ratio will fall at some point on the curve defined by the relative cost of energy and contacting equipment. The equipment changes necessary to lower reflux without altering product quality are the subject of several of the conservation options, e.g., tray retrofit, tighter control systems, and reduced pressure operation.

The same type of analysis can be made in designing heat-exchanger systems to recover heat from the products. Increasing the heat-exchange area will increase the heat recovery, but at a rapidly increasing cost. This again results in an optimum value, but this optimum will also change with a shift in the relative price of energy.

OPPORTUNITIES FOR ENERGY CONSERVATION

There are many possible options for conserving energy in distillation processes. In this report options are grouped according to the investment required:

- Options requiring little or no capital investment include improvements in maintenance and operating practices.

Fig. 4 Multiple flash theory of distillation.

- Options requiring moderate capital investment (normally authorized at the plant level) include upgrading the tower internals, adding insulation, and improving waste-heat recovery by adding heat-exchanger area.
- Options requiring major capital expenditures (normally requiring corporate approval) include heat pumping, addition of intermediate reboilers and condensers, advanced instrumentation and control systems, split-tower (or multieffect) distillation, and two-stage condensation.

The conservation options in the first two categories are equally applicable to both new and existing units. Some of the more sophisticated options, however, can be used economically only in new unit design. Options suitable for retrofit are stressed here because this area offers the greatest potential for improvement.

Since coverage of all possible conservation options to the extent described is beyond the scope of this document, those that appear most promising are presented in detail, and the others are briefly catalogued in a later section.

Each discussion is organized as follows:

- Description of the technique

Fig. 5 Schematic diagram of relationship between reflux ratio and number of plates or stages.

- Demonstration of the potential for energy conservation (usually by an example application)
- Guidelines for general application
- Calculation procedures to determine energy savings and the option cost

The plant engineer should go through the manual, option by option, first scanning the descriptions and example applications. Then he should examine the guidelines to determine whether or not an option is likely to be applicable to his distillation column. If it appears attractive, he should go through the calculation procedures to perform a preliminary economic analysis.

SUMMARY OF ECONOMICS FOR ENERGY CONSERVATION

The operating cost of a distillation column is composed of operating labor, utilities, maintenance labor, maintenance materials, and various capital-related charges. The utilities here are all fuel related, e.g., electricity, steam, cooling water, and the primary fuels. The cost of fuel-related items rose by more than 153% from 1972 to 1978, as compared with an increase of only 84% for labor-related charges.²

From Fig. 6, it is evident that the relative contribution of energy to the total operating cost is growing rapidly. The technology to reduce energy consumption is available, and it can give an acceptable rate of return on investment. Energy conservation is fast becoming a necessity if a plant is to remain competitive in the marketplace.

Here we concentrate on the effects of the historical price increase of energy during the 1970s, but there are strong indications that energy prices will continue to increase in the medium-term future. The factors tending to support this prediction include:

- Steady increases in energy consumption, not only in the United States but throughout the world
- Limited reserves of clean-burning fossil fuels
- Concentration of reserves in foreign lands, subject to arbitrary price control
- Conflict between energy efficiency and environmental concerns
- Apparent lag time before any alternate energy sources can make a significant contribution

Assuming that energy prices will continue to increase, an investment in energy conservation now will continue to grow more profitable.

Fig. 6 Fuel-cost component of overall operating cost for a refinery.

Chapter 2 Performance Evaluation of Existing Facilities

A thorough performance review is the first step when attempting to increase the profitability of a distillation operation. Rising energy costs have prompted closer evaluations of towers where operations have been adequate in the past. No matter how extensive the conservation program planned, a complete performance evaluation is the place to start. The purposes of this analysis are:

- Comparison of actual tower operations with design operations
- Accumulation of data required to build a model for optimization techniques
- Location of problem areas
- Establishment of a data base on which to plan future conservation programs

This evaluation can be broken into three steps:

- Gather design and historical operating data on the column, and prepare a test plan.
- Carry out the testing.
- Calculate various performance indicators for comparison with design performance and/or performance of other similar columns.

Since each tower or distillation system is unique, this is only a general outline to follow. The process engineering staff should shape an evaluation program to its own needs.

The procedure suggested in this chapter can require a significant expenditure of technical manpower. An engineer is needed to devise the test plan, calculate the various performance indicators, and, with the aid of a technician, collect field data and samples. A chemist or laboratory technician is needed for the analytical work. This manpower can be justified by the following benefits:

- Identification of immediate opportunities for energy savings and cost reduction
- Quick check of energy efficiency, which can indicate the need for and potential benefit of future energy-conservation measures
- Development of a data base from which more sophisticated conservation options can be evaluated
- Improved understanding of system performance

DATA REQUIREMENTS

Single-Column Binary Separation

All available printed information on the distillation tower, including process and instrumentation drawings, plot plans, design material and energy balances, and operational and maintenance records of the tower, should be found and studied. This material will provide information on what kind of data needs to be taken initially.

The flow diagrams should be inspected for points where temperatures, pressures, and samples of the fluids can be taken. Studying these diagrams will increase the engineer's understanding of the system. All pertinent information on the diagram should be verified in the field. Many revisions of operating hardware are never recorded on plant flow diagrams. Supplemental arrangements must be made for test purposes if adequate provisions for temperature, pressure, and flow measurement and for sample withdrawal are not currently available.

When all preparations for testing have been accomplished, the field work should be undertaken. The testing should be done when there are no unusual operating conditions and the unit is at steady state. Steady conditions should prevail throughout the test period, the length of which is determined by the amount and type of data and samples to be obtained. Normally the test should be long enough to collect all the data and obtain at least two sets of representative samples. Table 1 is a checklist of data

Table 1 Data Requirements for Single-Column Performance Evaluation

Flow rates	Temperatures (continued)
Feed	Overhead receiver
Overhead liquid product	Bottoms product
Overhead vapor product	Reboiler return
Reflux	Tower top
Bottoms product	Tower bottom
Reboiler circulation (if available)	Control tray (and any other available point in tower)
Reboiler heating medium	Cooling water (inlet—outlet)
Cooling water (if available)	Selected column insulation points
Pressures	Ambient
Overhead vapor	Reboiler heating medium
Overhead receiver	Steam, supply temperature
Feed at each preheater	Hot oil, inlet—outlet
Steam pressure	Fired reboiler, firebox
Intermediate trays (if possible)	Testing
Column bottom	American Petroleum Insti- tute gravity
Samples	American Society for Test- ing and Materials distil- lation
Feed	Light hydrocarbon compo- nent analysis (especially for key components)
Overhead liquid product	Other case specific param- eters
Bottoms product	Component analysis and vapor gravity
Overhead vapor	
Temperatures	
Feed at each preheater	
Overhead	

required for a binary distillation column. Figure 7 illustrates the points where physical measurements and samples should be taken.

In addition to data collected in the field, certain physical properties of the streams as they exist in the column must be determined. These include molecular weight, density, viscosity, latent heat of vaporization, specific heat, and vapor-liquid equilibrium constants.

Single-Column Multicomponent Separation

Most of the data acquired for binary separation also applies to multicomponent separation. Testing and data acquisition must be expanded to include side-stream products and circulating refluxes.

Distillation-Train Multicomponent Separation

Performance evaluation of an entire distillation train consists of an evaluation of each tower as described in the previous sections. Some streams will be common to adjacent columns, and a good test plan will provide for as little duplication of testing as possible. It is important, however, to note carefully the areas of interaction between the different towers. These are areas where significant energy savings are possible by such modifications as waste-heat recovery from bottoms-feed exchangers. It is important to obtain temperature profiles for all heat exchangers in the distillation train.

CALCULATIONS

The bulk of the information gathered before and during a field performance survey of operating distillation equipment can be processed into meaningful results by using appropriate chemical engineering principles. Since the main interest is energy conservation, the engineer will look for areas where calculations show that the tower is not operating up to design expectations and areas where higher energy prices have changed the optimum economic operating parameters.

Before calculating the tower's performance, we must establish the reliability of the test data. This is done by performing overall and individual component material and energy balances on the subject column. Examples of this type of calculation are given in many textbooks on separation processes.^{3-9,20,29}

The actual fitting of the data to a balance may show some discrepancies because of uncertainty in the precision and accuracy of the data. An error under 3% is considered excellent, but if the error approaches 10%, some or all of the data are unsatisfactory.¹⁰ Good engineering judgment and/or supplemental testing and measurement can compensate for any discrepancies. Obtaining good data before calculating detailed performance indicators is important.

Stage-to-stage calculations give the engineer the best picture of tower operation. This method solves for flow

rates and compositions of both liquid and vapor on each tray. By its very nature, this calculation is rigorous and time-consuming. It is recommended that the engineer use one of the readily accessible computer programs for modeling the distillation column. Most companies have programs of this sort available in-house. Smaller companies can purchase programs on the open market or contract computer services from a consultant. A recently published extensive listing of the available simulation packages¹¹⁻¹⁴ contained 29 comprehensive steady-state modeling programs and 34 distillation-specific programs.

From a computer simulation of tower operation, such parameters as reflux ratios, minimum reflux ratio, heat duties, internal vapor and liquid loads, and the number of theoretical plates can be determined. This provides the engineer with a great deal of information on which economic and technical evaluations can be made in the current study, as well as in any further studies.

Although the calculation procedures required to perform this simulation manually are tedious and expensive in terms of technical manpower, they are well documented and readily available. Any standard distillation text presents the details of many shortcut calculation procedures, as well as those for rigorous tray-to-tray calculations.³⁻⁹

PERFORMANCE INDICATORS

There are certain parameters that indicate how well a distillation column is using energy, and the performance calculations will determine these parameters. This manual discusses many options for energy conservation, and in each case the engineer can use factors determined by performance testing and calculations to aid in evaluating the options.

The best overall performance indicator is a quantity called the specific energy of distillation (Q^*), which was defined in a recent report as the energy required to make a distillation separation per pound of feed or products.¹ Values of the specific energy of distillation are listed in Tables 2 and 3 for columns of the petroleum and chemical industries. The value of Q^* can be calculated by dividing the reboiler duty by the feed rate. This provides the engineer a means to make a quick comparison of his column's energy consumption with that of the industry average for that type of column. Also listed in the tables are the number of trays, reboiler temperature, and condenser temperature, which can be used to characterize the average column. Chemical industry columns are listed in order of descending national energy consumption.

If a given column has a Q^* greater than the industry average, energy requirements can very probably be reduced significantly. It is still quite possible to effect energy savings even if the column is well below the average since the values given are averages of existing facilities and not a "state of the art."

Historical data can be used to evaluate instrumentation and control-system performance by plotting product qualities and specifications vs. time. The product-quality data should be collected on the shortest time interval available (a minimum of once per shift or once per day) to demonstrate the range of variation experienced by the

tower control system. A good control system should maintain product quality within $\pm 1\%$ of the specification.

The performance of tower internals is best indicated by the ratio of actual to minimum reflux. If the reflux ratio is greater than 1.3, there is a good chance for reducing energy requirements by retraying.

Table 2 Specific Energy of Distillation for Petroleum Refining and Natural Gas Liquids Columns*

Column description	Specific distillation energy consumption, Btu/lb charged	Key components	Number of trays	Reboiler temperature, °F	Condenser temperature, °F
Petroleum operations					
Crude distillation	193	Light naphtha/heavy naphtha, light distillate/heavy distillate, atmospheric gas oil/reduced crude	20–65	650–750	100–350
Vacuum distillation	132	Light vacuum gas oil/heavy vacuum gas oil/residue	9–14	700–800	150–250
Alkylation					
H ₂ SO ₄					
Deisobutanizer	510	Isobutane/ <i>n</i> -butane/isopentane	50	330	165
Deisopropanizer	510	Propane/isobutane	27	200	120
HF					
Deisobutanizer	436	Isobutane/ <i>n</i> -butane/isopentane	50	300–350	160
Depropanizer	382	<i>n</i> -Propane/isobutane	60	200	130
Propylene by-product	206	Propylene/propane	20	155	145
Rerun column	22	<i>n</i> -Butane/polymer	5	400	200
Catalytic hydrotreating and hydrorefining	101	Hydrogen sulfide, C ₁ –C ₄ /naphtha/furnace oil	20–40	300–600	150–330
Catalytic cracking fractionator	112	C ₁ –C ₄ /gasoline/light gas oil/heavy gas oil/slurry decant oil	30	675	150–300
Naphtha fractionator					
Catalytic cracking	132	<i>n</i> -Butane/isopentane	35	350–500	150–200
Crude atmospheric	132	<i>n</i> -Butane/isopentane	35	350–500	150–200
Thermal operations	132	<i>n</i> -Butane/isopentane	35	350–500	150–200
Catalytic hydrocracking	632	C ₁ –C ₄ /naphtha/diesel/bottoms	10–40	400–900	100–350
Catalytic reforming	132	C ₁ –C ₄ /gasoline	35	350–500	150–200
Sour-water strippers	240	Hydrogen sulfide/ammonia/water	40	260	240
Light-ends processing					
Depropanizer	365	Propane/isobutane	30–50	220–260	120–140
De-ethanizer	137	Ethane/propane	20–40	190	50
Iso/ <i>n</i> -butane splitter	153	Isobutane/ <i>n</i> -butane	60–80	130–120	100–180
Debutanizer	44	<i>n</i> -Butane/isopentane	30–50	300–450	150–200
Isomerization					
C ₄ feed deisobutanizer	1107	Isobutane/ <i>n</i> -butane	60–80	130–210	100–180
C ₅ feed deisopentanizer	664	Isopentane/ <i>n</i> -pentane	60–80	135–215	120–200
C ₅ + C ₆ feed deisopentanizer	664	Isopentane/ <i>n</i> -pentane	60–80	150–250	120–200
Natural gas processing					
De-ethanizer	265	Ethane/propane	30–50	120–150	0–20
Depropanizer	235	Propane/isobutane	30–50	220–240	120–140
Iso/ <i>n</i> -butane splitter	204	Isobutane/ <i>n</i> -butane	60–80	130–210	100–180
Debutanizer	105	<i>n</i> -Butane/isopentane	30–50	260–280	150–200
Iso/ <i>n</i> -pentane splitter	12	Isopentane/ <i>n</i> -pentane	60–80	145–225	115–195
Depentanizer	6	<i>n</i> -Pentane/hexane	30–50	300–350	250

*Data from Ref. 1.

Table 3 Specific Energy of Distillation for the Chemical Industry*

Process and column		Specific distillation energy consumption, Btu/lb of product†	Key components, light/heavy	Number of trays	Reboiler temperature, °F	Condenser temperature, °F
Sodium carbonate	Carbamate stripper	1,875	Ammonia/water		200	
Ethylene and propylene	Propylene	439	Propylene/propane	138	131	117
Styrene, from ethylbenzene	Styrene/ethylbenzene	1,946	Ethylbenzene/styrene	34	210	130
Urea	Ammonia still	1,334	Ammonia/water			
Liquid air	O ₂ /N ₂	158	N ₂ /O ₂	60/44	-289/-278	-309/-284
Ethylene and propylene	Absorber de-ethanizer	1,505	Ethane/propylene	30-50	200	50
Ethylene and propylene	Ethylene	295	Ethylene/ethane	73	14	-26
Ethylene and propylene	De-ethanizer	285	Ethane/propylene	40	146	14
Vinyl chloride, oxychlorina- tion process	High boiler	1,176	1,2 Dichloroethane/ 1,1,2 trichloroethane	30	250	185
Methanol, natural gas	Methanol	1,054	Methanol/water	60	280	200
BTX, sulfolane	Benzene	245	Benzene/toluene	53	264	180
<i>o</i> -Xylene	Xylene splitter	5,689	<i>m</i> -Xylene/ <i>o</i> -xylene	130	350	293
Acetic acid, acetaldehyde oxidation	Acid-acid finishing	2,362	Water/acetic acid	22	245	159
BTX, sulfolane	Extractive stripper	217	Heptane/benzene	15	375	275
Ethanol, direct hydration of ethylene	Stripper	4,576	Ethanol azeotrope/ water	17	240	205
Ethylene glycols, hydration of oxide	Evaporators	1,407	Water/ethylene glycol	15	255-360	221-353
BTX, sulfolane	Toluene	205	Toluene/xylene	45	313	235
Ethylene and propylene	Quench tower	248				
Butadiene, acetonitrile extractive distillation	Butadiene rerun	1,566	1,3 Butadiene/vinyl acetylene	130	120	100
Ethylbenzene, alkylation of benzene with ethylene	Benzene	874	Benzene/ethylbenzene	20	285	178
Ethanol, direct hydration of ethylene	Final	4,288	Ethanol azeotrope/ water	60	235	199
Ethylene and propylene	Demethanizer	116	Methane/ethylene	24	39	-130
Cumene, from benzene and propylene	Benzene	1,048	Benzene/cumene		330	176
Ethylene and propylene	Depropanizer	86	Propylene/C ₄	30	209	8
Dimethyl terephthalate, oxidation of <i>p</i> -xylene	Methanol purifier	1,008	Methanol/water	33	230	118
Acrylonitrile, Sohio	Carbonyl	1,579	Acetone/acrylonitrile	50	185	160
Adipic acid, air oxidation of cyclohexane	Concentration still	1,580		15	163	
Isopropanol, direct hydration of propylene	Drying column	1,343	Benzene/isopropanol	36	190	153
Terephthalic acid, oxidation of <i>p</i> -xylene	Acetic acid recovery	885			532	232
Acrylonitrile, Sohio	Acrylo/aceto splitter	1,401	Acrylonitrile/acetonitrile	70	220	200
Terephthalic acid, oxidation of <i>p</i> -xylene	Water removal	875	<i>p</i> -Xylene/water	42	275	212
Ethylene and propylene	Gasoline fractionator	352	C ₉ /fuel oil	35	373	189
Acrylonitrile, Sohio	Acrylo purifier	1,303	Acrylonitrile/cyanohydrin	20	212	210
Ethylene oxide, air	Refiner	1,127	Ethylene oxide/water	50	290	125
Vinyl acetate monomer, from ethylene and acetic acid	Separation and stripping	3,061	Vinyl acetate-water azeotrope/acetic acid- water azeotrope	35	257	162
Phenol, from cumene	Cumene	2,005	Cumene/phenol	38	226	115
Formaldehyde	Methanol	660	Formaldehyde/methanol	20	190	130
BTX, sulfolane	Xylene	141	Xylene/C ₉ aromatics	35	360	307
Butadiene, acetonitrile extractive distillation	Butenes extraction	1,095	<i>cis</i> 2 Butene/1,3 butadiene	75	185	110

(Table continues on following page.)

Table 3 (Continued)

Process and column		Specific distillation energy consumption, Btu/lb of product†	Key components, light/heavy	Number of trays	Reboiler temperature, °F	Condenser temperature, °F
Ethylbenzene, gasoline fractionation	Ethylbenzene	11,582	Ethylbenzene/ <i>p</i> -xylene	390	350	290
Adipic acid, air oxidation of cyclohexane	Cyclohexane	2,210		10		175
BTX, sulfolane	Solvent stripper	126	Sulfolane/C ₈ aromatics	35	375	170
Ethylbenzene, alkylation of benzene with ethylene	Ethylbenzene	560	Ethylbenzene/diethylbenzene	50	320	200
Ethylene glycols, hydration of oxide	Water removal	824	Water/ethylene glycol	16	333	152
Ethylene glycols, hydration of oxide	Ethylene glycol	507	Ethylene glycol/diethylene glycol	16	356	239
Phenol, from cumene	Heavy ends	826	Phenol/acetophenone	39	400	250
Vinyl acetate monomer, from ethylene and acetic acid	Refined vinyl acetate	1,209	Vinyl acetate/ethyl acetate	90	167	152
Phenol, from cumene	Phenol	798	Phenol/acetophenone	54	342	260
Methyl ethyl ketone, from <i>n</i> -butylene	Crude sec-butanol	3,392	sec-Butanol/water	68	228	205
Isopropanol, direct hydration of propylene	Azeotropic	942	Isopropanol/water	12	220	172
Acetaldehyde, direct oxidation of ethylene	Acetaldehyde still	988				
Styrene, from ethylbenzene	Ethylbenzene/benzene	240	Benzene/ethylbenzene	12	186	110
Hexamethylenediamine, hydrogenation of butadiene	Toluene	1,987	Toluene/3 & 4 pentenenitrile	28	220	140
Hexamethylenediamine, hydrogenation of butadiene	3 Pentenenitrile/adiponitrile splitter	1,947	3 Pentenenitrile/adiponitrile	50	230	150
Ethylbenzene, alkylation of benzene with ethylene	Tar removal	249	Benzene/diethylbenzene	12	392	275
Acrylonitrile, Sohio	Acetonitrile	941	Acetonitrile/water	60	220	193
Vinyl chloride, oxychlorination	Vinyl chloride, oxychlorination	244	Vinyl chloride monomer/1,2 dichloroethane	18	320	115
Vinyl chloride, oxychlorination	Finishing	244	Vinyl chloride monomer/1,3 butadiene	15	320	115
Styrene, from ethylbenzene	Styrene/tetraethylbenzene	219	Styrene/tetraethylbenzene	6	250	146
Vinyl chloride, oxychlorination	Crude dichloroethane	239	1,2 dichloroethane/1,1,2 trichloroethane	10	250	185
Isopropanol, direct hydration of propylene	Light ends	744	Diisopropyl ether/isopropanol—water	44	129	150
Propylene oxide, chlorohydrin process	Purification	678	Propylene oxide/propylene dichloride	15	180	125
Glycerine, using peracetic acid	Concentrator	8,897	Water/glycerine	4	310	287
Butadiene, acetonitrile extractive distillation	Butadiene stripper	354	1,3 butadiene/acetonitrile—water	20	285	105
Adipic acid, air oxidation of cyclohexane	Ketone/alcohol lower layer	710		13	230	204
Methyl ethyl ketone, from <i>n</i> -butylene	Light ends	2,078	sec-Butanol/water	62	216	183
Acetone, dehydration of isopropanol	Isopropanol	1,227	Isopropanol—water/mesityl oxide—water	52	212	177
Hexamethylenediamine, hydrogenation of butadiene	Methyl butene nitrile/3 pentenenitrile splitter	1,360	2 Methyl, 3 butenenitrile/3 & 4 pentenenitrile	90	212	140

(Table continues on following page.)

Table 3 (Continued)

Process and column		Specific distillation energy consumption, Btu/lb of product†	Key components, light/heavy	Number of trays	Reboiler temperature, °F	Condenser temperature, °F
Propylene oxide, chlorohydrin process	Oxide	529	Propylene oxide/propylene dichloride	50	200	68
Acetone, dehydration of isopropanol	Light ends stripper	1,051	Propene/acetone	61	177	120
Ethylene oxide, air	Desorber	197	Ethylene oxide/water	20	210	55
Acetic acid, acetaldehyde oxidation	Methyl acetate	310	Methyl acetate/water	44	224	133
Methanol, natural gas	Light ends	120	Dimethyl ether/methanol	60	275	115
Acetone, dehydration of isopropanol	Acetone	853	Acetone/water	50	210	135
Phenol, from cumene	1/2 Concentration	315	Cumene/dimethyl phenol carbanol		210	174
Methyl ethyl ketone, from <i>n</i> -butylene	Drying	1,347	Methyl ethyl ketone—water azeo/methyl ethyl ketone	62	194	169
Vinyl chloride, oxychlorination	Low boiler	114	Chloroprene/1,2 dichloroethane	77	200	160
Dimethyl terephthalate, oxidation of <i>p</i> -xylene	Dimethyl terephthalate purifier	228		30	487	388
Acetic anhydride, via ketene from acetic acid	Acetic acid	748	Acetic acid/acetic anhydride	101	267	252
Methyl ethyl ketone, from <i>n</i> -butylene	sec-Butanol purification	1,161	sec-Butanol/methyl heptanone	60	228	213
Vinyl chloride, oxychlorination	Hydrochloric acid	101	Hydrochloric acid/vinyl chloride monomer	19	200	-13
Dimethyl terephthalate, oxidation of <i>p</i> -xylene	Crude dimethyl terephthalate recovery	214		20	487	388
Phenol, from cumene	1/2 Crude acetone	252	Acetone/cumene	23	320	138
Vinyl acetate monomer, from ethylene and acetic acid	Vinyl acetate	368	Acetaldehyde/vinyl acetate	55	167	144
Cumene, from benzene and propylene	Cumene	200	Cumene/diisopropyl benzene		376	280
Cumene, from benzene and propylene	Propane	200	Propane/benzene		403	120
Acetone, from cumene	1/2 Concentration	502	Cumene/dimethyl phenyl carbinol			174
Isopropanol, direct hydrogenation of propylene	Finishing	287	Isopropanol/polymer	15	190	180
Methyl ethyl ketone, from <i>n</i> -butylene	sec-Butanol stripper	965	Water—sec-butanol—diisobutylene azeo/sec-butanol	15	230	205
Butadiene, acetonitrile extractive distillation	Propyne	141	Propyne/1,3 butadiene	40	120	100
Isopropanol, direct hydrogenation of propylene	Benzene recovery	263	Benzene/water	40	218	194
Acetic anhydride, via ketene from acetic acid	Dehydration	561	Water/acetic acid	40	252	200
Hexamethylenediamine, hydrogenation of butadiene	Adiponitrile	584	Adiponitrile/high boilers	36	400	369
Hexamethylenediamine, hydrogenation of butadiene	Methylglutaronitrile ethylsuccinonitrile	584	Adiponitrile/methylglutaronitrile	36	392	360
Hexamethylenediamine, hydrogenation of butadiene	<i>cis</i> 2 Pentenenitrile	584	<i>cis</i> 2 Pentenenitrile/3 & 4 pentenenitrile	31	190	140

(Table continues on following page.)

Table 3 (Continued)

Process and column		Specific distillation energy consumption, Btu/lb of product†	Key components, light/heavy	Number of trays	Reboiler temperature, °F	Condenser temperature, °F
Acetone, from cumene	1/2 Crude acetone	402	Acetone/cumene	23		136
Acetic acid, acetaldehyde oxidation	Acetaldehyde recovery	170	Acetaldehyde/methyl acetate	34	324	135
Acetone, from cumene	Refined acetone	383	Acetone/water	20	207	136
Glycerine, using peracetic acid	Initial heavy ends	2,721	Diisobutylketone/heavy impurities	33		300
Adipic acid, oxidation of cyclohexane	Ketone/alcohol stripper	237	Cyclohexane/water	13	230	204
<i>o</i> -Xylene	Heavy ends	331	<i>o</i> -Xylene/heavy ends	36	327	295
Isopropanol, direct hydrogenation of propylene	Heavy ends	203	Isopropanol—water azeo/polymer	20	219	177
Ethylbenzene, alkylation of benzene with ethylene	Benzene drying	601	Water/benzene		180	176
Acetic anhydride, acetaldehyde oxidation	Acetic anhydride	424	Acetic anhydride/ethylidene diacetate	32	350	284
Acetic anhydride, acetaldehyde oxidation	Acetic acid	384	Acetic acid/acetic anhydride	50	285	245
Hexamethylenediamine, hydrogenation of butadiene	Hexamethylenediamine	409	Adiponitrile/hexamethylenediamine	40	400	206
Phenol, from cumene	Dehydration	136		11	275	125
Dimethyl terephthalate, oxidation of <i>p</i> -xylene	Methyl <i>p</i> -toluate recovery	111		42	439	327
Hexamethylenediamine, hydrogenation of butadiene	Medium boiler concentrator	348	Medium boilers/hexamethylenediamine	40	350	307
Methyl ethyl ketone, from <i>n</i> -butylene	Heavy ends	482	Methyl ethyl ketone/sec-butanol	42	217	110
Vinyl chloride, oxychlorination	Ethylene dichloride	43	1,2 Dichloroethane/1,1,2 trichloroethane	10	250	185
Acetic anhydride, acetaldehyde oxidation	Dehydration	305	Water/acetic acid	50	180–293	133–212
Hexamethylenediamine, hydrogenation of butadiene	Medium boiler stripper	300	Medium boilers/hexamethylenediamine	40	333	248
Styrene, from ethylbenzene	Toluene/ethylbenzene	535	Toluene/ethylbenzene	28	280	230
Cyclohexane, benzene hydrogenation	Stabilizer	98				
Acrylonitrile, Sohio	Hydrogen cyanide	134	Hydrogen cyanide/water	15	284	166
Adipic acid, oxidation of cyclohexane	MBA stripper	126				200
Vinyl chloride, oxychlorination	Degassing	30	Hydrochloric acid/vinyl chloride	15	250	115
Ethanol, direct hydration of ethylene	Light ends	146	Acetaldehyde/ethanol—water azeo	44	244	198
Acetic anhydride, via ketene from acetic acid	Acetic anhydride	213	Acetic acid/dehydroacetic acid	8	366	267
Styrene, from ethylbenzene	Benzene/toluene	26	Benzene/toluene	34	248	176
Ethylene glycols, hydration of oxide	Diethylene glycol	42	Diethylene glycol/triethylene glycol	12	428	309
Ethylbenzene, alkylation of benzene with ethylene	Diethylbenzene	26	Diethylbenzene/triethylbenzene	10	392	275
Acetaldehyde, direct oxidation of ethylene	Degasser	90				
Acetic anhydride, via ketene from acetic acid	Light ends	175	Acetic acid/acetic anhydride	68	265	228

(Table continues on following page.)

Table 3 (Continued)

Process and column		Specific distillation energy consumption, Btu/lb of product†	Key components, light/heavy	Number of trays	Reboiler temperature, °F	Condenser temperature, °F
Glycerine, using peracetic acid	Flasher	846	Water/glycerine		347	238
Acetone, from cumene	Light ends	103	Acetaldehyde/acetone	42	147	115
Acetic acid, acetaldehyde oxidation	Ethyl acetate	43	Ethyl acetate/water	4	212	159
Acrylonitrile, Sohio	Cyanohydrin stripper	67	Acrylonitrile/cyanohydrins	8	230	
Acetic anhydride, via ketene from acetic acid	Acetic acid heavy ends	128	Acetic acid/acetic anhydride	55	264	252
Glycerine, using peracetic acid	Splitter	750	Acetic acid/diisobutyl ketone	24	230	143

*Data from Ref. 1.

†Generally based on the primary (or larger) product.

If the relative volatility (α) of the key components is less than 1.1 or if one of the two streams separated represents less than 5% of the feed, it may be worthwhile to investigate alternative separation methods.

For a new unit design with a reboiler duty greater than 17 million Btu/hr, split-tower distillation should be seriously considered. If the column also has a low differential temperature, heat pumping may be the best solution. For very large columns with overhead temperatures high enough for attractive heat recovery, two-stage condensation should be considered.

If a column exhibits nonideal behavior, the use of intermediate reboilers and/or condensers may be attractive. This is also true of columns where use of a heat pump is considered but the differential temperature is too high.

Options such as insulation and waste-heat recovery should be uniformly studied. Any surface with a skin temperature more than 30°F different from ambient should have an optimum insulation thickness calculated. Any stream going to artificial cooling should be studied for the possibility of waste-heat recovery.

Chapter 3 Energy Conservation in Operation and Maintenance Practices

The preceding chapter outlined a data acquisition and analysis procedure for evaluating the performance of an existing unit. Performance indicators were listed from which the energy efficiency of each column could be judged against other similar columns in the industry. If a given column exceeds the specific energy requirement for its class, it can almost certainly benefit from a close examination of operating and maintenance practices. Even if it is below the specific energy requirement, it can possibly benefit from this analysis.

This chapter discusses moderate energy savings (up to about 10%), but the costs involved are very low. To

evaluate major capital expenditures for energy conservation accurately, we must first establish a baseline energy requirement indicative of present equipment working at its optimum efficiency.

IMPROVEMENTS IN OPERATING PRACTICES

Several process variables should be examined when attempting to reduce energy consumption on an existing column (see Fig. 8). The most obvious is the reflux. The reflux ratio is one of the prime indicators of efficiency in the operation of a distillation column. Any change in reflux

Fig. 8 Important process variables.

will cause a change in the product quality unless it is balanced by another change in operations. Several variables may be manipulated to allow a reduction in reflux and, thus, a reduction in energy consumption.

One set of variables that should be examined is the product specifications. It is common practice to set the unit target specifications slightly higher than the ultimate specifications for sale or further processing. This minimizes production of off-test product during normal operating swings. A careful study of typical product quality, as compared with the ultimate specifications, should be made to determine whether the current operation is overly conservative. If a product is collected in tankage and sold by the batch, it is possible to "test up" the batch to compensate for some periods of being off specifications. Bringing product quality down to the specified limit can result in significant energy savings.

Along the same line, the ultimate specifications should be critically examined. The consequences of off-test product must be determined and balanced economically against the potential energy savings. If product specifications are absolute limits, off-test product cannot be sold, creates a dangerous situation, or causes severe economic penalties. Most product specifications are not absolute, but after years of operation they may seem to be. A critical reappraisal of the specifications may be warranted as energy prices rise.

For example, let us examine a tower making a single split of two components. The lighter material (component A) is 20% of the feed and the heavier material (component B) is 80%. Components A and B have a relative volatility of 2.0, and the tower is being operated at a reflux ratio of 1.3. Under these conditions a relaxation of product specifications on the overhead stream from 99 to 98% component A and on the bottom stream from 98 to 96% component B can cut energy consumption by more than 10%.

Feed-plate location is another important factor. The feed should enter the tower at the point where the composition on the tray matches the composition of the feed. If it enters anywhere else, a loss of efficiency caused by blending will occur. An improper ratio of stripping trays to rectifying trays and, thus, deficient operation in one section can also result. Most columns are designed with several optional feed points, but these are rarely used. In many cases feed-plate relocation can be implemented with a minimum capital expenditure and can produce significant energy savings. Tray-to-tray calculations (usually performed by a computer simulation package) are required to determine the concentration profile, which in turn determines the optimum feed tray.

For example, assume that a study of an existing column (by direct sampling or by simulation) determined that the feed is entering the column two actual trays too high. The distillation system is characterized by a relative volatility of 1.5, a reflux ratio of 1.3, a separation factor of 10,000, and

a 10% concentration of the lighter component in the feed. The separation factor is a mathematical expression that quantifies the degree of separation: $S = y(1 - x)/x(1 - y)$, where y is the overhead vapor composition and x is the bottoms liquid composition. For this system, correcting the feed location can cut energy requirements by 1.7%. This can often be done with no capital investment since most towers have several alternate feed points.

A study of proper feed location should also consider the feed condition (liquid, gas, or mixed phase). The physical state of the feed can have a large effect on the tower heat balance, shifting duty from the reboiler to reflux with increased gas in the feed. This can also unbalance the number of trays in stripping and rectifying service, resulting in inefficient operation. Any plans to increase feed preheat should include a check of the feed condition at the tower entry.

Separation by distillation is easier at lower pressures for most materials. The term that describes this is relative volatility (α). The relative volatility of most pairs of components increases with decreasing pressure. Therefore, lower pressures can be used to achieve the same separation at lower reflux ratios.

The operating pressure of most columns is set to allow condensation of the overhead vapors with cooling water (or air). Since this limits the overhead temperature to 100 to 125°F, the pressure has to be quite high to condense some of the more volatile components. Even under this constraint, ample opportunity exists for reducing column pressure when the cooling capacity is high (at night and during cold or rainy weather). Sophisticated control systems for such pressure reduction are available, but a similar reduction can also be made by an alert operator who has been taught to look for the opportunity.

It is also worthwhile to collect data on the composition of the overhead vapors. This can be used in a flash calculation to determine the pressure required to condense the overhead at the condenser temperature. The existing operating targets may possibly have been set too conservatively in times of low-priced energy.

Estimating the energy saving potential of process optimization is very difficult since it depends on the initial process condition. The low cost of such a study makes it worthwhile, however, even if energy savings are minimal. In addition, the data gathered in this phase is often useful in evaluating other energy-conservation options.

The procedures used to calculate the energy savings resulting from relaxation of product specification or reduction in pressure are covered in detail in a later section on advanced instrumentation and control (see Chap. 5). A correlation is made between separation factor, relative volatility, theoretical trays, and the ratio of boilup to feed. For changes in specifications, calculate the change in the separation factor and relate this value to the change in boilup. For reduced pressure operation, determine the

relative volatilities at each pressure and relate this change to a reduction in boilup.

Changes in feed-plate location are much more difficult to assess. Determination of the optimum tray requires computer simulation. This computer output should include calculations of reboiler and condenser duty. The energy savings can then be determined by a comparison of the base case and the optimum feed-point case.

IMPROVEMENTS IN MAINTENANCE PRACTICES

The proper combination of preventive maintenance, corrective maintenance, cleaning, and inspection of equipment can effect a significant energy savings. Maintenance and cleaning of heat exchangers and fractionating trays, as well as general equipment inspection, are especially important to energy-efficient operation of distillation systems.

Process-stream heat exchange is one of the primary methods of conserving energy in distillation. Its effectiveness can be greatly reduced by heat-exchanger fouling, however. Fouling can be corrected by frequent cleaning, but that may require a unit shutdown unless the exchanger can be bypassed.

Preventive measures to reduce heat-exchanger fouling include:

- Treatment of process stream (desalting, blanketed storage tanks, filtering, etc.)
- Proper equipment design (use of baffles to increase shell-side velocity)
- Use of proper metallurgy to control corrosion fouling
- Use of antifoulants

An antifoulant is a chemical that prevents either the formation of solids or the adherence of solids to the heat-exchanger surface. Antifoulants are usually effective at low concentrations and are becoming more economically attractive as fuel costs continue to rise.

A good example of the use of antifoulants is their application to the crude preheat-exchanger train in petroleum refining. A significant reduction in the fouling rate is achieved when an antifoulant is used, as evidenced by a slower decay of heat-exchanger efficiency (Fig. 9). The annual energy savings is 3.2 billion Btu per thousand barrels of crude per operating day (BPOD) at an antifoulant cost of \$1290 per year per 10^3 BPOD. This represents a net annual savings of more than \$3000/ 10^3 BPOD of crude, with energy valued conservatively at \$1.40/ 10^6 Btu.¹⁵

The basis for a good heat-exchanger maintenance program is continuous monitoring of heat-exchanger performance. Heat-transfer data should be taken at a minimum interval of once a month, preferably much more often. These data will aid in identifying exchangers that have fouling problems and provide a data base from which to calculate potential savings. Vendors of industrial chemicals will then be able to suggest antifoulants and/or corrosion

inhibitors that will fit the physical and chemical properties of the system.

The condition of the trays or packing inside a distillation column is very important to efficient operation. Plugging, fouling, and corrosion of the trays can upset process conditions and require more energy to achieve the same degree of fractionation. Operating outside design conditions, e.g., flooding, also degrades efficiency. These problems can be detected most easily by a tray-to-tray pressure-drop survey. If individual tray pressure taps are not available, some indication of tray condition can be gained by examining the overall tower pressure drop.

The distillation system should be regularly inspected to identify any sources of energy loss. Any type of leak—

Fig. 9 Heat-transfer improvements with the use of an antifoulant. (a) Outlet temperature decay rates. (b) Crude unit heat duty decay rates.

steam, hydrocarbon, or cooling water—means a loss of energy. Regular visual inspections of equipment for missing or damaged insulation can be supplemented by occasional infrared surveys to pinpoint hot spots. The steam system should be carefully checked for leaks and for proper functioning of steam condensate traps.

These may seem like small items, but they are not insignificant. A $\frac{1}{8}$ -in.-diameter leak in a 150-psig steam line will lose 110,000 Btu/hr, which represents a cost of \$2200/year if energy is valued at \$2.50/10⁶ Btu. Viewed cumulatively, such small losses represent a substantial cost item.

Chapter 4 Moderate Investments to Improve Energy Efficiency

This section describes energy-conservation options that require low-to-moderate capital investments. Many of these projects require expenditures small enough to be approved at the plant level although the expense may vary from case to case. All the options described here are equally applicable to new designs and to retrofit situations. The options in this category include:

- Insulation
- Waste-heat recovery
- Retraying for higher efficiency and lower pressure drop

These options require investments ranging from several thousand dollars to about \$50,000 and can reduce energy requirements by as much as 20%, which results in payout times of under 2 years at an energy cost of \$2.50/10⁶ Btu.

INSULATION

The optimum thickness for insulation is a function of the cost of energy lost from a bare surface and the cost of insulating the surface. This relationship is shown graphically in Fig. 10. The many specific factors that determine these costs have been defined comprehensively by the Thermal

Fig. 10 Relationship determining optimum insulation thickness.

Insulation Manufacturers Association (TIMA) in their publication *ECON-I* and by the Department of Energy in *ETI-Economic Thickness for Industrial Insulation*.¹⁶

Since the optimum thickness of insulation increases as energy costs rise, it is recommended that the insulation thickness be set according to the projected price of fuel at the half-life of the insulation. This precludes the rapid obsolescence of a given insulation thickness and the recurring high cost that comes from adding insulation one layer at a time. If plant fuel-cost projections failed to anticipate the recent rapid escalation in fuel prices or if insulation thickness was set according to contemporary fuel prices, additional insulation can probably be justified.

Most plants have enough insulation on the process piping, but valves, flanges, and manways are often left bare. The irregular shape of these fittings increases the cost of conventional insulation, and there is also a significant possibility of the insulation's being ruined by leakage. Both of these problems can be overcome by using removable—reusable insulating covers, which are available off-the-shelf in a range of standard sizes. The final cost is relatively low because the covers can be installed quickly by unskilled labor. The insulating blanket material is impervious to product leakage and weather.

Energy-Conservation Potential

Insulating covers for process fittings are discussed here to illustrate the economics of adding insulation. Figures 11 and 12 show the energy saved and the payout time (for fuel valued at \$2.50/10⁶ Btu) for adding removable insulating covers to process valves and flanges. Similar results can also be achieved on manways. As would be expected, the payout time decreases with increasing skin temperature and line size, but attractive payouts are indicated even at the low end of the range (300°F and 4-in. line size). Thus the economically attractive range of application is wide.¹⁷

Guidelines for Application

Every distillation column, piping, and fitting is a reasonable candidate for insulation. The farther the operating temperature is from atmospheric, the more economically attractive the application. This, of course, includes refrigerated surfaces as well as high-temperature surfaces.

The once accepted rule of thumb that insulation cannot be justified for temperatures between ambient and 200°F should no longer be taken for granted. Depending on the size and geometry of the surface, the ambient conditions,

and the magnitude of the heat loss, insulation can often be attractive well below 200°F.

Calculation Procedures

Optimum Insulation Thickness. The equation for determining the optimum insulation thickness for a flat surface¹⁸ is:

$$t = \left[\frac{(k)(\Delta T)(K_v)}{dC/dt} \right]^{1/2} - \frac{k}{h}$$

where k = thermal conductivity of the insulation [(Btu/in.)/(hr/ft²/°F)]

ΔT = temperature difference between process and ambient (°F)

K_v = value of energy conserved (\$/Btu)

$\frac{dC}{dt}$ = (cost of insulation including setup charges)/(area x thickness x payout time for capital x on-stream time) [\$/ft² in. hr]

h = heat-transfer coefficient from the insulation surface to the surroundings [Btu/(hr/ft²/°F)]

For a pipe, the equation for optimum insulation is¹⁸:

$$r_i \ln \frac{r_i}{r_p} + \frac{k}{h} = \left[\frac{\left(1 - \frac{k}{r_i h}\right) (k)(T)(K_v)}{dC/dt} \right]^{1/2}$$

where k , h , ΔT , K_v , and dC/dt are as defined in the equation for a flat surface, r_i is the outer radius of insulated surface (in.), and r_p is the outer radius of uninsulated pipe (in.).

This same type of analysis (with slightly different factor definitions) was performed by TIMA, and the results were

Fig. 11 Economics for insulating valve bodies with removable-reusable covers. (Data from Ref. 17.)

published in tabular form in *ECON-I*. This publication is available on request, as is the DOE publication *ETI-Economic Thickness for Industrial Insulation*. A FORTRAN computer program has also been written to perform this analysis. Member companies of TIMA will run this program for a specific plant, or the program itself can be purchased for in-house analysis.

Economics for Insulating Fittings. To estimate losses from any fitting, we must determine the following factors:

- Exposed area
- Bare surface temperature
- Ambient conditions (temperature, wind, rain, etc.)
- Thermal conductance of the insulation

The exposed areas of various standard-sized flanges and valves are given in Table 4. On the basis of the bare surface temperature and assumed ambient conditions, the hot-surface heat-loss factor was calculated (Table 5). The heat-loss factor from an insulated surface is shown in Table 6 (Ref. 17) for a thermal conductance of 0.27 Btu/(hr/ft²/°F) at 300°F mean temperature and 0.32 Btu/(hr/ft²/°F) at 400°F.

The energy savings for insulating a fitting is then expressed by:

$$\begin{aligned} \text{Energy savings} &= \text{Loss from bare surface} - \text{loss from insulated surface} \\ &= (\text{Bare area}) \times (\text{Hot-surface loss factor}) - \\ &\quad (\text{Insulated area}) \times (\text{Insulated-surface loss factor}) \end{aligned}$$

Fig. 12 Economics for insulating flanges with removable-reusable covers. (Data from Ref. 17.)

The value of energy savings can be determined from the plant's fuel cost or steam cost, whichever is appropriate.

The cost of removable—reusable insulating covers as of Dec. 1, 1978, is shown in Table 7. The resulting payout time for each of the standard sizes was previously shown for a fuel value of \$2.50/10⁶ Btu in Figs. 11 and 12.

WASTE-HEAT RECOVERY

Here we study the means for more fully utilizing the heat in each stream leaving a distillation column. The recovery of this heat is more efficiently accomplished by heat exchange with other process streams which may be related to the same column, another nearby column, or a separate process that could economically use additional heat.

Two areas should be considered for improvement studies—existing exchangers with relatively large temperature approaches and process streams with no waste-heat recovery. An existing exchanger for process heat recovery may already be in use, but if the heat-exchange area were greater more waste heat would be recovered. There may also be streams not currently used for heat exchange whose temperatures are sufficiently high to heat another process stream, preheat combustion air to a fired heater, or even generate steam. Perhaps the economics were not favorable for such an arrangement when the unit was designed, but the situation should be reexamined. In studying this option, the cost of heat-exchange area must be balanced against the savings realized from heat recovery.

Table 4 Exposed Areas of Fittings*

Pipe size, in.	Flanges		Valve/flange factor†
	Area, ft ²	Equivalent pipe length, ft	
3	1.48	1.62	2.55
4	1.91	1.62	2.62
6	2.78	1.60	2.73
8	3.77	1.67	2.79
10	5.20	1.85	2.82
12	6.71	2.04	2.82

*Data from Ref. 17.

†Valve area = flange area x valve/flange factor.

Table 5 Hot-Surface Heat-Loss Factors*

Ambient conditions	Surface temperature, °F	Loss factor, Btu/(hr/ft ²)
70° F, natural convection only	300	654
	400	1145
	500	1806
	600	2666

*Data from Ref. 17.

Energy-Conservation Potential

An analysis of waste-heat recovery for a naphtha debutanizer feed—bottoms exchanger is presented here. Figure 13 shows the existing exchanger, with the new exchanger being added in series. The area of the original exchanger is approximately 1640 ft², and the feed outlet temperature is 250° F.

Calculations were made for various feed outlet temperatures to determine the amount of additional area required. The savings were then determined from the increased heat recovery, which in turn reduced reboiler duty. A payout period can be calculated from the plot in Fig. 13 by dividing the exchanger cost by the annual savings.

The upper limit of feed temperature increase is the bottoms product temperature. In Fig. 14 the annualized cost of the recovery operation (cost of energy lost with any given size of new exchanger plus installed cost for new exchanger) is plotted against additional exchanger area. The minimum point on this curve is the optimum heat-exchange area. Two different energy costs were used in calculating the optimum point, and it is clear that more area can now be justified with higher energy prices. There may be certain constraints on such a calculation, and in this specific case the feed-column inlet condition was the limit on additional area. For this example, the payout is approximately 0.6 years.

In this example a new exchanger was added in series to increase the heat-transfer area. Another alternative is the use of extended surface tubes in existing exchanger shells.

Table 6 Insulated-Surface Heat-Loss Factors*

Hot-surface temperature, °F	Heat-loss factor, Btu/(hr/ft ²)		
	70° F, natural convection	55° F, natural convection	55° F, light wind†
300	56.4	59.4	60.7
400	87.1	89.9	91.4
500	121.6	124.4	126.7
600	160.5	163.5	166.0

*Data from Ref. 17.

†Average of 1 mph.

Table 7 Typical Costs of
Insulated Fitting Covers*

Pipe size, in.	Valve cover	Flange cover
2	\$62	\$21
4	\$83	\$30
8	\$123	\$46
12	\$180	\$64

*Data from Ref. 17.

For boiling heat-exchanger applications, a porous boiling surface can be applied to the tubes to increase heat-transfer coefficients by a factor 10 to 30 times that for a bare tube. Spiral exchangers and plate-and-frame exchangers are also attractive when space is limited.

Time for Savings to Equal Investment
at \$2.50/10⁶ Btu

Feed temperature, °F	Payout, year
270	0.67
280	0.61
290	0.60
300	0.59
310	0.61
320	0.62
330	0.64

Fig. 13 Economics for an additional feed-bottoms exchanger for a naphtha debutanizer.

Guidelines for Application

All process streams should be considered candidates for process heat recovery. A survey of all streams leaving the distillation column should be performed to identify those which are most attractive for waste-heat recovery. Any product stream that is cooled by cooling water or air should be considered as a possible heat donor stream. An extensive search in the general area for a stream that could benefit from additional heat must then be performed. A quick calculation of possible heat transfer should be made, and the cost of equipment to effect the transfer should be calculated. If the resulting economics are reasonably good, a more thorough investigation is warranted.

Because the economics for each heat-recovery situation are slightly different, it is difficult to generalize about temperature guidelines above which heat recovery should be considered. As a general rule, any stream that is currently being cooled artificially (cooling water, air coolers, or refrigeration) can be considered for heat recovery.

Calculation Procedures

Two cases are of interest in calculating the economics of waste-heat recovery—streams having no process heat recovery.

Fig. 14 Graphic analysis of the optimum heat-exchanger area.

ery and those having insufficient process heat recovery. Here we calculate the economics of increased heat recovery by adding a second exchanger in series with an existing exchanger. For streams with no existing heat exchange, a similar procedure should be followed, with a value of zero used in the existing exchanger area of the analysis.

The first step is to gather all pertinent design and historical operating information on the existing exchanger, including fouling data and any loss of area because of plugged tubes.

The data should be verified and augmented by field testing. The temperature and pressure in and out of each side of the exchanger should be measured. Samples of both streams should be taken to determine the pertinent physical properties.

The following calculations should then be performed:

1. Calculate the heat transferred, Q , on each side of the exchanger for sensible heat gain or loss:

$$Q_1 = MC_p \Delta T_1$$

where M is the mass flow rate (lb/hr), C_p is the heat capacity [Btu/(lb/°F)], and ΔT is the differential temperature across the exchanger for the subject stream.

2. Calculate heat transfer for evaporation or condensing service:

$$Q_1 = MC_p \Delta T_1 + (FE/C)(\Delta H_{vap})(M)$$

where FE/C is the fraction evaporated or condensed and ΔH_{vap} is the latent heat of vaporization (Btu/lb).

3. Compare the heat transferred on the hot side, $(Q_{hs})_1$, with that on the cold side, $(Q_{cs})_1$. There should be no more than a $\pm 5\%$ difference between the two. If that limit is exceeded, examine the data to resolve the difference.

4. On the basis of the value for heat duty, calculate the heat-transfer coefficient:

$$U = \frac{Q_1}{A_1 (LMTD)_1}$$

where A is the exchanger area (ft²), U is the overall heat-transfer coefficient [Btu/(hr/ft²/°F)], and $LMTD$ is the log mean temperature difference (°F), which is defined by

$$LMTD_1 = \frac{(T_{hi_1} - T_{co_1}) - (T_{ho_1} - T_{ci_1})}{\ln \left[\frac{T_{hi_1} - T_{co_1}}{T_{ho_1} - T_{ci_1}} \right]}$$

where T_{hi} = hot-side inlet temperature (°F)

T_{ho} = hot-side outlet temperature (°F)

T_{ci} = cold-side inlet temperature (°F)

T_{co} = cold-side outlet temperature (°F)

5. Assume that a second exchanger is added in series to the existing exchanger and that it will have the same U value (overall heat-transfer coefficient). Make calculations over the pair of exchangers as a single unit. Increment the

cold-side outlet temperature, $(t_{co})_2$, by 20°F and recalculate the cold side heat duty, $(Q_{cs})_2$. Since $(Q_{cs})_2 = (Q_{hs})_2$, recalculate the hot-side outlet temperature, $(T_{ho})_1$, based on the new value for $(Q_{hs})_2$. Recalculate the $LMTD$ based on the new temperature profile. Solve for the area:

$$A_2 = \frac{Q_2}{U(LMTD)_2}$$

Then the incremental area of the second exchanger is:

$$\Delta A = A_2 - A_1$$

The approximate cost of this exchanger can be determined by standard plant-design and cost-estimating texts, by in-house procedures, or by vendor estimates. A figure of \$40 to \$60 per square foot can be used as a preliminary value for carbon steel exchangers. The incremental energy recovery is:

$$\Delta Q = Q_2 - Q_1$$

Determine the effective price of energy for the plant and calculate the dollar value of this energy savings. Calculate a simple payout time to act as a preliminary economic screening indicator:

$$\text{Payout} = \frac{\text{Investment (\$)}}{\text{Savings (\$/year)}}$$

6. Increment the cold-side outlet temperature by another 20°F and repeat the process. Continue this iterative procedure until the payout time increases beyond an attractive level or until physical constraints are met. Physical constraints may be the ultimate convergence of the stream temperatures ($T_{co} \rightarrow T_{hi}$ or $T_{ho} \rightarrow T_{ci}$) or external limits based on downstream processing.

7. Select the case with the greatest process heat recovery which shows an acceptable payout time for a more detailed economic analysis.

TRAY RETROFIT

In designing a new distillation column, we must determine the optimum number of trays to make the desired separation. The more theoretical stages present, the less reflux is required to make any given separation. A balance must then be achieved between the cost of more trays and the energy savings possible from lowering the reflux. The bounding values are minimum trays at total reflux and minimum reflux with an infinite number of trays (see Fig. 5).

The actual economic optimum is determined by the price of contacting equipment relative to that of energy. Figure 15 shows the results of such an economic study on a debutanizer in an ethylene plant.¹⁹ Note that the optimum reflux ratio (here defined as actual reflux divided by minimum reflux) decreased from 1.4 in 1960 to 1.15 for the present. Continued operation at a reflux ratio of 1.4 is

shown to incur approximately \$60,000 net annual increased operating expense.

Most existing distillation columns were designed under much different economic conditions than the present.¹³ The purpose of this section is to present shortcut methods to calculate the energy savings attributable to any given increase in the number of theoretical trays. This calculation can then be performed for several points, tray costs can be estimated, and the optimum reflux ratio (or optimum number of theoretical trays) can be determined.

The number of theoretical trays in a distillation column is the product of the number of actual trays and the tray efficiency. Therefore, an increase in theoretical stages can be achieved by:

- Increasing the total number of actual trays by enlarging the column or reducing the tray spacing
- Replacing the existing trays with a more efficient type

Increasing the number of trays by enlarging the fractionating vessel can be prohibitively expensive, but it is the most reliable way to be sure of increasing the total number of theoretical stages. Reducing the tray spacing to increase the number of trays in an existing vessel also has several drawbacks and could be quite expensive because of the labor-intensive modification of tray supports. Lowering the tray spacing could also cause a loss of efficiency from entrainment and a resultant loss in capacity. The cost estimates presented here are based on completely retraying the column with a more efficient contacting device.

Many types of contacting devices are available. They can be divided into broad classes, such as trays (bubble-cap, valve, and sieve) and packings (rings, saddles, and spirals). The efficiency and the pressure drop are the critical properties of the contacting device which determine the potential for energy conservation.

It is very difficult to generalize about tray efficiency (or height equivalent of a theoretical tray for packing) because of its dependence on the specific application. Consequently, no attempt will be made here to identify specific trays or packings that might result in improved efficiency. An analysis is presented to determine the energy savings possible at various levels of increased tray efficiency. This information can be combined with rough cost data to make a preliminary economic analysis. If the result looks attractive, tray and packing vendors can be approached to determine the maximum feasible increase in efficiency and a more precise cost.

Energy-Conservation Potential

As an example of the economics of tray retrofit, an analysis was made for a naphtha debutanizer. This type of column is common in refineries, natural gas processing plants, and light olefin units. This is a relatively easy split characterized by low reflux ratios. The payouts for this kind of operation are not as attractive as those for a column splitting close boiling components with high reflux ratios. This case was chosen to illustrate the fact that reasonable payouts can be achieved for upgrading the trays of almost any reboiled distillation column.

Figure 16 shows the debutanizer and its basic operating conditions. Energy savings are calculated for an increase in theoretical trays of 2, 5, and 10%. The cost of replacement for 35 actual trays was estimated to be \$11,300 for the equipment and between \$5650 and \$11,300 for installation (depending on site-specific factors). The payout time for this investment is shown in Fig. 17 for two different energy prices and two installation costs. This demonstrates that tray retrofit on the debutanizer becomes attractive if a 3 to 4% increase in theoretical trays can be achieved. For a column with 35 actual trays, this represents only a 2.2% increase in tray efficiency.

Guidelines for Application

Theoretically, energy savings can be achieved by increasing the number of theoretical trays in any reboiled distillation column. Columns in which all the required heat enters with the feed (e.g., crude distillation columns and catalytic cracker main fractionators) cannot achieve this type of energy reduction.

The amount of savings for each case depends on how far the existing conditions are from the optimum balance of reflux to theoretical stages. The prime indicator of this is

Fig. 15 Comparison of annual cost vs. reflux ratio for a debutanizer in a large ethylene plant for 1960 and predicted 1975-1980 energy costs. (Data from Ref. 19.)

Retrofit case	1	2	3
Increase theoretical trays by L/D (Gilliland)	2%	5%	10%
Energy savings, 10 ⁶ Btu/hr	0.12	0.32	0.59
Annual savings at \$2.50/10 ⁶ Btu	\$2,376	\$6,369	\$11,682
Payout for \$16,950, years	7.1	2.7	1.5
Payout for \$22,600, years	9.5	3.5	1.9

Cost of tray retrofit

Equipment \$11,300
 Installation \$5,650 to \$11,300
 Total \$16,950 to \$22,600

Fig. 16 Example for analysis of economics of tray retrofit for a naphtha debutanizer with the following operating conditions: feed, 20,000 BPOD unstabilized naphtha; distillate product, 2200 BPOD (11% of feed); bottoms product, 17,800 BPOD (89% feed); light key, normal butane; heavy key, isopentane; reflux, 4800 BPOD; L/D = 2.16; (L/D)_{min} = 1.44 (Underwood); actual-to-minimum reflux ratio, 1.5; actual trays, 35; theoretical trays, 21.6 (Gilliland); tray efficiency, 61.7%; minimum theoretical trays, 12.0 (Fenske); condenser duty, 8.5 x 10⁶ Btu/hr; and reboiler duty, 35.0 x 10⁶ Btu/hr.

the ratio of actual to minimum reflux, where the following guidelines apply:

Ratio of actual to minimum reflux	Potential for cost reduction
>1.3	Excellent chance for improvement
1.3 to 1.1	Marginal but deserves further study
<1.1	Improvement doubtful

These are general guidelines, and there are many exceptions. A very energy intensive process (such as cryogenic separation of air) can justify reflux substantially lower than 1.1 times the minimum. On the other hand, if extensive heat recovery is coupled with low-cost energy sources (such as low-pressure steam), operation with reflux as high as 1.5 times the minimum can be economically attractive.

The option to upgrade the trays in a distillation column should always be seriously considered when making other large changes, especially if the vessel is to be modified. Trays must also be replaced periodically because of corrosion or accidental damage. Upgrading the trays (rather than replacement in kind) can offer an excellent payout in this situation because only the differential cost between tray types must be justified.

Fig. 17 Economics for retraying a naphtha debutanizer.

Calculation Procedures

Energy Savings—Method 1. The most accurate way to estimate the energy savings from an increase in theoretical stages is by computer simulation. If the operation of a specific column was simulated under the performance evaluation, all that remains is to run several cases specifying various efficiencies but maintaining the same separation. The required reflux and the corresponding reboiler heat duty for each case will be output.

Energy Savings—Method 2. If a process simulation package is not available, the energy savings can be estimated by a more generalized procedure. A recent study prepared by the Merix Corporation¹ derived the following general relationship:

$$\frac{d \ln Q}{d \ln N} = \frac{-2.3524(N)(N_{\min} + 1) \left[\frac{R - R_{\min}}{R + 1} \right]^{0.4332} (R + 1)}{(N + 1)^2 (R_{\min} + 1)}$$

where Q = reboiler heat duty (Btu/hr)

N = number of theoretical stages

$N_{\min}$ = minimum number of theoretical stages

R = reflux ratio (reflux/distillate)

$R_{\min}$ = the minimum reflux ratio

For small finite variations in N (up to about 10%), the calculated value can be used in the following relationship:

$$\frac{\Delta Q}{Q} = \left[\frac{d \ln Q}{d \ln N} \right] \left(\frac{\Delta N}{N} \right) \text{ or } \Delta Q = \left[\frac{d \ln Q}{d \ln N} \right] \left(\frac{\Delta N}{N} \right) (Q)$$

One of the assumptions used in the derivation of this equation is that reboiler duty equals condenser duty. If the duties are not equal for a specific application, using the smaller duty for Q will give a more realistic estimate of the energy savings.

The Merix report went one step further and used a computer simulation package to define N , $N_{\min}$, and $R_{\min}$ from the following input parameters:

1. Mole fraction of light key in the feed
2. Relative volatility (α)
3. Separation factor
4. Ratio of actual to minimum reflux

The results of this program are presented in tabular form in Volume II of that report. The resulting four-dimensional data array is too voluminous to be included here, but one page is shown as an example in Fig. 18.

Energy Savings—Method 3. It is also possible to hand calculate the estimated energy savings by the following methods, as described in any standard distillation text:³⁻⁹

1. Calculate minimum reflux by one of the following shortcut methods: Underwood, Brown—Martin, or Colburn.
2. Calculate the minimum number of theoretical plates at total reflux by the Fenske method.
3. Calculate the number of theoretical plates for the existing reflux ratio by the Gilliland correlation.
4. Increment the number of theoretical plates by some percentage, and back calculate through the Gilliland correlation to the new reflux ratio.

Fig. 18 Effects of increasing the number of trays and decreasing the separation factor on the condenser duties of distillation columns.¹ Mole fraction of the light key in the feed = 0.40. Distribution coefficient (α) = 1.20.

5. Calculate the energy savings from reduced reflux by:

$$\Delta Q = [(\Delta L)(\Delta H_{\text{vap}})] + [(\Delta L)(C_p)(\Delta T)]$$

(latent heat + sensible heat)

where ΔL = change in reflux

ΔH_{vap} = latent heat of vaporization

C_p = heat capacity

ΔT = temperature differential between the reflux accumulator

The Gilliland correlation is most often presented in graphical form, but an equation was developed to model the curve by Eduljee³⁰:

$$\text{Let } Y = \frac{N - N_{\text{min}}}{N + 1}$$

$$\text{and } X = \frac{R - R_{\text{min}}}{R + 1}$$

$$\text{then } Y = 0.75 (1 - X^{0.5668})$$

$$\text{or } X = \left(1 - \frac{Y}{0.75}\right)^{1.76}$$

The equation form makes it much easier to evaluate the effect of small changes in the number of theoretical plates in step 4.

Cost of Trays—Method 1. The most accurate method for determining tray costs is to get preliminary bids from a tray manufacturer. Most large tray vendors will provide this service at no charge in the hope of promoting business.

Cost of Trays—Method 2. A rough order-of-magnitude estimate of tray costs can be determined from Table 8. These estimates are based on vendor costs for three different tray types, in two different metallurgies and three different order volumes. Prices are given per square foot of tray area and can, thus, be scaled to estimate roughly the cost of most common trays. These values should not be extended to very large columns or to columns using multipass trays.

Table 8 Cost of Replacement Trays (\$/ft²)*

	1 Tray		20 Trays		50 Trays	
	Carbon steel	Stainless steel	Carbon steel	Stainless steel	Carbon steel	Stainless steel
Valve trays	66	76	12	15	9	12
Bubble cap	112	130	21	28	15	20
Sieve trays	60	68	10	14	8	11

*1978 values.

Cost of Tray Installation

Tray installation costs depend on site-specific factors. Normally installation costs range from 50 to 100% of the cost of the trays themselves.

Chapter 5 Energy Conservation Requiring Major Capital Investment

Although impressive energy savings can be achieved in the low-investment range, the greatest savings potentials are in options requiring larger investments. At current energy prices, even the large expenditures can easily be justified. Some of these options can be retrofit to an existing unit, but others require such extensive modifications that they are usually applicable only to new designs. The energy-conservation options in this category include:

- Upgrading instrumentation and control systems
- Heat pumping
- Intermediate reboilers and condensers
- Two-stage condensation
- Split-tower (or multieffect) distillation

These energy-conservation projects require investments ranging up to 3 million dollars, but they can reduce energy requirements by as much as 90%. At an energy price of \$2.50/10⁶ Btu, these investments can be paid out in 1 to 3 years.

ADVANCED INSTRUMENTATION AND CONTROL SYSTEMS

The control system for a distillation column has two basic functions—to stabilize and to optimize the opera-

tion. When applied properly, the control system should ensure the production of on-test products at minimum cost within the bounds of other equipment constraints.

Conventional distillation control systems are designed only for stabilization, relying on operator intervention for any optimization. Advanced control systems that use direct material-balance control, feedforward control, on-stream product analyzers, and digital computers for optimization are available. Such systems can give more stable operation, control the column at the optimum operating point within equipment constraints, reduce the need for operator intervention, and reduce energy requirements and operating costs.

A conventional control system for a simple two-component fractionator is shown in Fig. 19. Product flows are level controlled. The reflux is set independently at a fixed rate, which is usually conservatively high. Product quality is controlled inferentially by varying the reboiler heat input to hold a constant temperature on some critical tray. This type of system is susceptible to variations in the heat balance, which cause swings in product quality.

The effect of poor product-quality control on energy requirements is shown graphically in Fig. 20. Product-

Fig. 19 Distillation column with conventional control.

quality variations may average out, a period of overfractionation balancing a period of underfractionation, and result in a product with the desired specifications. There is an energy penalty associated with this fluctuating operation, however, since the energy requirement is not a linear function of the separation factor. Energy can be saved by exercising tighter control on product quality (reducing overfractionation).

Direct Material-Balance Control

Direct material-balance-control systems, illustrated in Figs. 21 and 22, can reduce swings in product quality, typically to within $\pm 0.5\%$ of the specification. The improvement is achieved by using a control variable that is coupled more directly to product quality; that control variable is product flow rate. Consider the case of a simple two-component separation. The amount of each component entering the tower is fixed by the feed rate and the relative component concentration. The purity of each product stream is then determined by the relative rate at which each product is withdrawn. With an onstream analyzer giving a continuous indication of product quality and resetting the product flows to meet specifications, very tight control can be achieved.²¹

Another feature incorporated into many advanced control systems is feedforward control. A conventional control system using feedback control must first sense a disturbance in one of the controlled parameters and then take corrective action to return the column to steady state. A

feedforward system senses factors that could disturb the operation and compensates for them before the disturbance occurs. A classic example is the use of an analyzer to sense changes in the feed composition and adjust tower operations accordingly. Ideally, feedforward control could achieve perfectly stable operations, whereas feedback control can only minimize the magnitude of disturbances. A combination of feedforward and feedback control can achieve similar results with much simpler process models.

Direct material-balance control can be used with analog-type controllers or with digital (computer) systems. Much of the savings that result from reducing overfractionation can be achieved with relatively low-priced analog equipment. The addition of a computer further expands control-system capabilities to both stabilize and optimize the operation. A conventional control system attempts to maintain steady-state conditions, and, within those bounds, it can be very effective when properly designed. A computer control system can go beyond steady-state control by adjusting unit operations to an optimum within the bounds of the equipment design and safety constraints.

Energy Conservation Potential for Direct Material-Balance Digital Control

A direct material-balance digital control system was retrofitted to a series of light hydrocarbon fractionating columns in a gas processing plant. The direct material-balance system was used to decrease overfractionation, and

Fig. 20 Effect of poor control on energy requirements.

Fig. 21 Direct material-balance control by using the distillate to control product quality.

Fig. 22 Direct material-balance control by using the bottoms product to control composition.

optimizing algorithms were used to increase heat-transfer efficiency in the hot oil reboiler system. This project was justified on the basis of a 10% decrease in energy requirements, but it achieved a better than 20% decrease in actual operation. The actual savings were about 35 million Btu/hr for the 22,000-BPOD plant, or a savings of nearly \$750,000/year with fuel valued at \$2.50/10⁶ Btu.^{2,2}

Guidelines for Applying Direct Material-Balance Control

Similar savings from a decrease in the amount of overfractionation can be achieved on any *reboiled* distillation column (but not on columns where all the heat enters with the feed). The exact amount of the savings for any case depends on the efficiency of the conventional control system. Accumulated performance data showing product quality compared with the specification should be examined. If the product quality does not stay within $\pm 1\%$ of the specification, an economic improvement can probably be achieved with an advanced control system. Ideally the product quality should be expressed in terms of the desired separation of the light key and heavy key rather than some inferred quality, such as flash point, viscosity, or color. These characteristics may be too insensitive to the separation to detect small changes.

Calculation Procedures

The energy saved by reducing overfractionation can be estimated by a relationship developed by Shinskey^{2,3} and refined by R. C. Seeman and A. E. Nisenfeld. Define the generalized separation function, Y :

$$Y = \frac{\log S}{N \log \alpha}$$

where N is the number of theoretical trays, α is the relative volatility, and the separation factor, S :

$$S = \frac{y(1-x)}{x(1-y)}$$

where y is overhead vapor composition and x is bottoms liquid composition. Define the generalized boilup function, X :

$$X = \frac{V\alpha^{1.68}}{FN^{0.32}}$$

where V is the vapor rate or boilup and F is the feed rate. Then X and Y are related as follows:

$$Y = 0.286X \quad \text{for } X < 1.825$$

$$Y = \frac{1}{[8.35/(X^2 + 1.163)]^{1/2}} \quad \text{for } X > 1.825$$

or

$$X = \frac{Y}{0.286} \quad \text{for } Y < 0.522$$

$$X = \left(\frac{8.35}{1/Y^2 - 1.163} \right)^{1/2} \quad \text{for } Y > 0.522$$

To calculate potential energy savings, first calculate the amount of boilup required to achieve a separation that exactly meets product specifications. This represents the minimum energy required to make that separation with the given column equipment. Then calculate the boilup required to make the separation indicated by historical product quality. The difference between these two figures is the potential reduction in boilup (or energy consumption at the reboiler). A latent heat of vaporization must be applied to convert the moles of reduced vapor boilup to an energy basis.

This correlation was developed for an even split between overhead and bottoms products. For a column with an imbalanced split, a more accurate estimate of the energy savings can be arrived at by determining $(V/F)_1$ for present operations, $(V/F)_2$ for proposed operations, and the percentage change:

$$\frac{(V/F)_1 - (V/F)_2}{(V/F)_1} \times 100$$

and applying the percentage to whichever is smaller, the reboiler or condenser heat duty.

The cost of an advanced control system is highly dependent on case-specific factors. It may range from as low as \$10,000 for a small analog control system to several hundred thousand dollars for a large digital control system. The only reliable method of estimating the cost of an advanced control system is to solicit vendor estimates.

Advanced Control of Crude Distillation Units

The atmospheric crude distillation column is the largest single consumer of energy in distillation. It offers significant opportunities for conservation because of its size and complexity. An advanced control system using a digital computer for optimization can cut crude distillation energy requirements by 8 to 10% and will usually pay itself out in less than 1 year. Similar results can be achieved with other large multiproduct fractionators.

An advanced control system can reduce crude distillation energy requirements by:

- Reducing the amount of overflash to a minimum
- Optimizing circulating reflux flow rates
- Optimizing crude flow rates through parallel exchanger trains and heater passes
- Optimizing the furnace combustion efficiency by maintaining proper excess air for varying fuels

Proper crude unit-design practice calls for 2 to 4% overflash, which condenses to provide an internal reflux for the stripping trays below the feed point. With a conventional control system, it is very difficult to determine the degree of overflash for any given set of operating conditions since overflash is an internal phenomenon. This difficulty often results in conservative operation with as much as 8 to 10% overflash. An advanced control system with a digital computer can monitor the degree of overflash by making a continuous series of heat and material balances. Overflash can then be minimized by controlling the withdrawal rate of the lowest side-stream product or by adjusting the heater outlet temperature. Overall energy requirements of most crude distillation units can be cut by several percent simply by keeping overflash in the design range.

An advanced control system can also improve performance in the upper sections of the column. Traditional operating philosophy calls for maximizing one product (which varies with seasonal demand) while keeping all products on-test. With a conventional control system, such operation often results in overloading one section of the column, thereby reducing tray efficiency, increasing the overlap between streams, and increasing steam and/or heat requirements in the side-stream stripper. A digital control system can overcome this by monitoring the internal vapor-liquid loadings and maximizing the desired product within design loading constraints. It can manipulate the

circulating reflux flows to both improve fractionation and optimize heat transfer. This type of control can reduce energy requirements by about another 2%.

A digital control system can also be used to improve heat-transfer efficiency in fired heaters and exchanger trains where the crude is divided into a number of parallel passes. With a conventional control system, the operator usually tries to maintain equal pass outlet temperatures. This procedure can approach optimum operation, but that is seldom the case in actual practice. Computer algorithms can monitor the heat transfer to each pass and adjust the flow rates to optimize the total heat transfer, within safe operating limits. Such dynamic control could be accomplished with an analog system, but the safety factor and on-line time would be much lower. A digital system can boost heater (or heat-exchanger) efficiency enough to reduce fuel consumption by 0.75 to 3%.

A computer can also be used to improve the combustion efficiency of the heater by controlling the amount of excess air. Analog control is adequate if a single fuel with a relatively stable heating value is used. However, gas shortages and industrial curtailments are encouraging the use of multiple fuels, such as a combination of refinery gas, natural gas, and fuel oil. When such fuels are used, direct correlation of stack oxygen readings with the excess combustion air is no longer valid, as can be seen in Fig. 23. Computer control of the excess air can improve the

Fig. 23 Relationship between stack O₂ and excess air for multifueled heaters.

efficiency of natural draft heaters with damper air controls by 0.5 to 1.0% and by up to 4% for systems with better air measurement and control.²¹

Guidelines for Application

The techniques described for the atmospheric tower of a crude distillation unit should be applicable to any large column performing a multicomponent separation with all heat energy entering with the feed. Many of these techniques (such as optimizing heat transfer in parallel passes and combustion control) are also applicable to other types of distillation.

Calculation Procedures

The energy savings, on a percent basis, for each factor of advanced control of a crude unit were given in the discussion. An estimate of the savings applicable to a specific unit can be made by applying that percentage to the atmospheric column heater duty.

The comments on system cost for direct material-balance control apply here also. The range of costs for advanced digital control of a crude column is probably \$100,000 to \$1,000,000. A typical cost for control hardware, process interface, system design, and start-up is about \$365,000. A cost quote tailored to a specific system can be obtained from an instrument vendor.

Floating-Pressure Control

Floating pressure control is another possible energy-conserving control system. For most hydrocarbons the relative volatility (or ease of separation) increases with decreasing pressure. Floating pressure control takes advantage of this property by maintaining the tower pressure at the lowest value at which the overhead vapors will condense. The pressure is, therefore, "floating" on the condenser capacity. Significant energy savings theoretically can result from taking advantage of increased cooling capacity at night, during the winter, and during rain, especially on columns equipped with air-cooled condensers.

Although this system has a significant potential for energy conservation, it has had limited acceptance. A difficulty in sensing column constraints is one problem. The exotic and unfamiliar equipment required (such as pressure-compensated temperature sensors and differential vapor-pressure sensors) is another. This technology is available, but significant development work is needed on each specific application to tailor the control system to the appropriate constraints.

The potential energy savings of reduced-pressure operation can be accurately determined by simulation or can be estimated by:

1. Determining the relative volatility (α) at the current pressure and the proposed pressure

2. Calculating the amount of boilup required to make the desired separation with each relative volatility by applying the relationship given in the direct material-balance calculations

The costs of floating pressure control are even harder to estimate than those for other advanced control systems. Reasonable cost estimates can be generated only by working closely with instrumentation vendors or in-house instrumentation engineers.

Summary

Advanced control systems for distillation are among the most promising energy-conservation options. They are almost universally applicable and are easily retrofitted to existing units, often without the need for a unit shutdown. They can make significant energy savings possible and, because of the improved data they provide, can sometimes point out other conservation opportunities. The cost of these systems varies widely with the size and complexity of the application, and it is, therefore, difficult to generalize. Advanced control systems can usually be paid out in about 1 year on energy savings alone, but they also offer a significant potential for improved profitability by increasing throughput and/or product yields.

HEAT PUMPING

In conventional distillation, heat is added to the column at the reboiler (usually from steam or a fired heater) and rejected at the overhead condenser (usually to cooling water or a refrigerant). The overall energy requirements of the column can be reduced significantly by using a heat pump for integrating these two functions.

A basic heat-pump cycle using an external refrigerant is shown in Fig. 24. Mechanical compression is used to raise the temperature level of the refrigerant above that of the reboiler. The heat liberated by the condensing refrigerant then reboils the bottoms liquid. The liquid refrigerant leaving the reboiler is evaporated in the overhead vapor condenser, and the cycle continues. Thus the heat rejected from the overhead is used to reboil the bottoms, with the aid of added mechanical work by the compressor.

In two variations on the basic heat pump, one of the column fluids acts as the refrigerant. Vapor recompression (Fig. 25) uses the overhead stream as the working fluid, thereby eliminating the need for an overhead condenser. Reboiler flashing (Fig. 26) uses the bottoms stream as the working fluid, thus eliminating the reboiler.

Heat-pump economics are very case specific. In many cases the heat pump cannot be applied for technical reasons, and in other cases it offers a poor to marginal payout. In applications for which it is well suited, however, the heat pump can achieve significant savings and a rapid payout. In some new designs, use of a heat pump can

reduce the initial investment as well as the operating expenses.

Energy-Conservation Potential

A propane-propylene splitter was chosen to illustrate heat-pump economics. Table 9 shows the operating and economic factors for a conventional splitter, a heat-pump retrofit, and a heat pump incorporated into a new design. Note that the column conditions were not varied in any of these cases; this permits the evaluation of the heat pump as an external energy-conservation option. Application of the heat pump reduced energy consumption by 88%, but it shifted the energy source from low-pressure to high-pressure steam. The resulting payouts were 1.2 years for the

new design and 2.1 years for the retrofit. The retrofit case assumed no salvage values for the existing reboiler and condenser. As the salvage values of these exchangers approach market value, the retrofit payout would approach that of the new design.

Both heat-pump applications assumed that no internal changes would be made to the conventional design. Heat pumping makes it possible to operate at a much lower pressure; this was constrained in the conventional case by condensation of the overhead stream with cooling water. Lowering the pressure increases the relative volatility, thus allowing a reduction in the reflux ratio and/or the number of trays. When coupled with other conservation options (such as high-efficiency trays and enhanced surface exchangers), this can result in a capital expenditure well below that for the conventional design and in much lower operating costs.^{24,25}

Guidelines for Application

Although heat pumping can produce significant energy savings, its range of applicability is limited. Some general guidelines on its application are:

- Compression work is determined primarily by the differential temperature (ΔT = reboiler temperature — condenser temperature). Heat pumping is best suited to columns with a low ΔT , ideally 65°F or less. This is the most important criterion, and it would override any of the others in case of conflict.
- Heat pumping can be most easily justified in a new unit design where column operating conditions can be modified to achieve an economic optimum.
- For retrofit situations, the following conditions apply:
 1. Columns currently using a refrigerated overhead condenser are prime candidates.

Fig. 24 Heat pumping in distillation with an external refrigerant.

Fig. 25 Vapor recompression.

Fig. 26 Reboiler flashing.

- Columns with an overhead temperature between 100 and 250°F show a marginal payout because they can use relatively inexpensive utilities (cooling water and low-pressure steam).
 - Columns using high-pressure steam in the reboiler with no heat recovery at the condenser are good candidates.
 - Columns using heat recovery at the overhead condenser show a marginal to poor payout.
- If vapor recompression or reboiler flashing is used, the process stream must be a suitable refrigerant. Any fluid can be considered suitable if

$$\frac{\Delta H_{\text{vap}}}{C_p T_2} \approx E_c$$

where ΔH_{vap} = latent heat of vaporization (Btu/lb)

C_p = heat capacity [Btu/(lb/°F)]

T_2 = compressor discharge temperature (°R)

E_c = compressor efficiency^{2,5}

- If both the overhead stream and the bottoms stream are suitable refrigerants, vapor recompression is the favored route for low-pressure columns, and reboiler flashing is favored at high pressure.

Calculation Procedures

The following steps outline the calculations necessary to estimate the energy savings and major equipment sizes for a heat pump using reboiler flashing:

- Refer to a pressure–enthalpy (Mollier) diagram for the appropriate working fluid (see Fig. 27). Locate the initial conditions—saturated liquid at the column bottom temperature and pressure.

2. The fluid is cooled by flashing across the expansion valve, which can be represented by an adiabatic expansion to the condenser temperature. Note enthalpy H_1 .

3. The fluid is then vaporized at constant temperature and pressure in the column overhead condenser. This step is represented by a horizontal line to the saturated vapor line.

Fig. 27 Generalized Mollier diagram for reboiler flashing.

Table 9 Economics for the Application of a Heat Pump to a Propane–Propylene Splitter*

Operating and economic factors	Conventional tower design	Reboiler flashing as a retrofit	Reboiler flashing on a new column
Column pressure, psia	275	275	275
Overhead temperature, °F	115	115	115
Bottoms temperature, °F	135	135	135
Steam consumption, 10 ⁶ Btu/hr	101†	12‡	12‡
Cooling water, 10 ⁶ Btu/hr	101	6	6
Utility costs, 10 ⁶ dollars/year	3.06	1.10	1.10
Increased maintenance, 10 ⁶ dollars/year		0.28	0.28
Savings, 10 ⁶ dollars/year		1.68	1.68
Additional capital cost, 10 ⁶ dollars/year		3.44	2.04
Simple payout, years		2.1	1.2
After-tax ROI, %		24.4	41.2

*All costs are in 1978 dollars.

†Low-pressure steam (valued at \$2.50/10³ lb).

‡High-pressure steam (valued at \$3.50/10³ lb).

Note enthalpy H_2 . The available condenser duty is then given by

$$Q_{CA} = M(H_2 - H_1)$$

where Q_{CA} is the available condenser duty (Btu/hr) and M is the reboiler mass flow rate (lb/hr). This value also represents a savings in cooling water from the conventional case. The heat duty for a trim condenser (if necessary) is given by

$$Q_{CT} = Q_C - Q_{CA}$$

where Q_{CT} is the trim condenser duty and Q_C is the total condenser duty (conventional case).

4. The next processing step is compression of the fluid vapors back to the column bottom pressure. This is represented on the diagram by an isentropic (constant entropy) compression line. Note enthalpy H_3 . The ideal work of compression is defined by

$$W_c = M(H_3 - H_2)$$

where W_c is the ideal work of compression. The compressor power is then defined by

$$P_c = \frac{W_c}{E_c} = \frac{M(H_3 - H_2)}{E_c}$$

where P_c is the compressor power and E_c is the compressor efficiency.

5. The conventional reboiler would have vaporized the column bottoms at constant pressure to the intersection of the saturated vapor line. Note enthalpy H_4 . The conventional reboiler duty is then given by

$$Q_{RC} = M(H_4 - H_1)$$

where Q_{RC} is the conventional reboiler duty. The energy savings attributable to the application of a heat pump is then given by

$$\begin{aligned} \Delta Q_{HP} &= Q_{RC} - P_c = M(H_4 - H_1) - \frac{M(H_3 - H_2)}{E_c} \\ &= M \left[(H_4 - H_1) - \frac{(H_3 - H_2)}{E_c} \right] \end{aligned}$$

These calculations give the basic data needed to size the overhead condenser, trim condenser, and compressor for a reboiler flashing system. Cost estimates for this equipment can be determined by using standard plant-design texts, in-house procedures, or vendor estimates.

The calculation procedures for vapor recompression and the external refrigerant cycle are similar to those for

reboiler flashing. More detailed procedures for all three versions can be found in sources listed in the bibliography.

INTERMEDIATE REBOILERS AND CONDENSERS

Conventional distillation is thermodynamically both expensive and inefficient because heat is added at its highest (most valuable) level and removed at its lowest (most difficult to recover) level. This situation can be eased by the use of intermediate reboilers and condensers, as illustrated in Fig. 28. Intermediate reboilers and condensers can reduce energy costs by substitution of a less expensive form

Fig. 28 Distillation column with intermediate reboiler and condenser.

of energy, increased opportunity for heat recovery, and improved internal operation of the column.

An intermediate reboiler can often use less expensive energy than a conventional reboiler, e.g., low-pressure instead of high-pressure steam. Although this results in a reduction in operating cost, it does not achieve any direct savings of energy. However, since many plants have an excess of low-pressure steam that must be vented if not used, its use in an intermediate reboiler could be considered an energy savings.

The same analysis can be applied to the condenser system. If the conventional column uses an expensive refrigerated condenser, significant savings could result from using an intermediate condenser with cooling water to reduce the refrigeration load.

A lower level of heat input and a higher level of heat removal also increase the opportunities for heat recovery by process heat exchange. Since the differential temperature between the intermediate reboiler and condenser is small compared with the overall column ΔT , the range of applicability of the heat pump principle is expanded greatly. This type of operation is illustrated in Fig. 29.

Fig. 29 Heat pumping with intermediate reboiler and condenser.

The operation of a fractionating column can be depicted graphically by the classic McCabe–Thiele diagram, shown in Fig. 30(a). A conventional column has an operating line broken into two sections, the slopes of which represent the internal reflux in the stripping and rectifying sections of the column. The addition of intermediate reboilers and condensers results in a four-part operating line that more closely follows the equilibrium line, as shown in Fig. 30(b). This can result in improved fractionation and lower energy consumption, but the benefits are minimal for systems exhibiting ideal behavior.¹

Figure 31 shows a McCabe–Thiele diagram exhibiting nonideal behavior in the form of a “pinch” zone in the

Fig. 30 McCabe–Thiele diagram for binary distillation with intermediate heating and intermediate cooling. (a) Conventional column operation. (b) Column with an intermediate reboiler and a condenser. Between x_1 and x_2 use lower operating lines.

stripping section. Since the operating line cannot cross the equilibrium line, a high reflux is required in the stripping section. This requirement can be reduced substantially by the addition of an intermediate heater, as shown by the modified operating line.

Energy-Conservation Potential

Very little documented data could be found on the specific economics of intermediate reboilers and condensers. A very careful analysis should be made of each specific application, including a rigorous fractionation simulation. This option was included primarily because of its interaction with other attractive conservation options (such as heat pumping).

Guidelines for Application

The following are general guidelines for the application of intermediate reboilers and condensers:

- This technology is applicable generally only to new designs since significant changes in the internal vapor-liquid loadings occur. It could be considered for retrofit if done concurrent with other major changes, such as retraying.
- Columns exhibiting a high degree of nonideality (such as azeotropes) should be given strong consideration.
- Columns where other process streams are available to provide heating and/or cooling at the intermediate levels may benefit from intermediate reboilers and condensers.

Fig. 31 McCabe-Thiele diagram for system with pinch region.

- Columns with large ΔT , which use expensive forms of energy (such as high-pressure steam and refrigeration), should be considered.
- Columns where use of a heat pump is being considered but the overall ΔT is too large are potentially attractive.

Calculation Procedures

The only way to accurately estimate the impact of intermediate reboilers and condensers on distillation energy consumption is by rigorous tray-to-tray calculations. These calculations are normally done by a process-simulation package. The first step is to model the existing (or proposed conventional design) column. Then shift part of the reboiler duty and condenser duty to intermediate trays and rerun the case until the same separation is achieved. Examine the energy requirements and operating parameters. Vary the draw trays and heat duties of the intermediate exchangers between runs to seek an optimum.

TWO-STAGE CONDENSATION

Much of the energy input into a distillation column is eventually removed at the overhead condenser. This heat energy is lost to the atmosphere if cooling water or air is used as the cooling medium. It is standard practice to recover as much of this energy as possible by process heat exchange, but a significant portion is lost because of the low temperature required in the distillate accumulator.

Two-stage condensation is a novel adaptation that will increase the energy available for process heat exchange. Energy is saved by condensing the vapors in two stages, thus preventing subcooling of the reflux. The first stage condenses only enough liquid to reflux the tower. The second stage condenses and cools the distillate product.

Energy-Conservation Potential

A conventional overhead condensing system of an atmospheric crude column is shown in Fig. 32(a). All the overhead vapors are cooled and condensed by process heat exchange down to some lower limit for economical heat recovery. The balance of the condensing-cooling is done by air coolers. This results in 40 million Btu/hr process heat recovery and 60 million Btu/hr loss to the atmosphere.

A two-stage condensation system for the same column is shown in Fig. 32(b). Only the reflux is condensed in the first stage, and the resulting reflux accumulator temperature is the same as that of the conventional tower top. The vapors leaving the first stage are then condensed in a conventional manner. All the latent heat of the reflux is recovered, along with a considerable portion of that for the distillate product. The result is 70 million Btu/hr process heat recovery and only 30 million Btu/hr loss; this represents a 75% increase in recovered heat.²⁶

The conditions at the first-stage separator are identical to those at the conventional tower top. Thus the first stage

acts like an additional theoretical fractionating stage. This can have a beneficial effect on product quality and recovery and can also provide additional energy savings if applied to a reboiled tower (see Chap. 4, Tray Retrofit). No credit should be assumed for any effects of better fractionation in the example crude tower.

Since the cost of this type of condenser system modification is relatively high, it can be most easily justified for large towers. The estimated cost for the example crude atmospheric tower is about \$376,000. The energy savings would amount to about \$600,000/year if energy is valued at \$2.50/10⁶ Btu, resulting in a simple payout of 0.6 year.

	Case 1	Case 2
Heat recovered, 10 ⁶ Btu/hr	40	70
Heat lost, 10 ⁶ Btu/hr	60	30
Energy savings, 10 ⁶ Btu/hr		30
Savings, dollars/year		594,000
Incremental investment, dollars		376,000
Simple payout, years		0.6

Fig. 32 Economics of two-stage condensation. (a) Case 1, conventional condensation. (b) Case 2, two-stage condensation.

Guidelines for Application

The two-stage condensation system is applicable when the following guidelines are met:

- There is a large spread between the bubble point and dew point of the overhead stream. This is typical of wide-boiling-point mixtures commonly encountered in petroleum refining.
- Overhead temperature is above the limit for economical process heat recovery.
- The technology is most easily justified on large units where economy of scale comes into effect.

Generalized Calculation Procedure

The amount of energy saved can be calculated easily. The energy savings for a two-stage system over conventional condensation are realized from the fraction of the total overhead for reflux times the heat lost to the atmosphere in the conventional case. The following equation is used to calculate the savings:

$$Q_s = Q_a \left(\frac{L}{L + D} \right)$$

where Q_s = savings (Btu/hr)

Q_a = heat duty of air coolers, conventional case (Btu/hr)

L = reflux rate

D = distillate rate

These savings are compared with the cost of retrofitting such a system. This is usually a major capital investment, and a significant study should be performed to estimate the cost of the project. Preliminary costs can be obtained from published cost correlations or from vendors. The necessary equipment to make the modification includes a new reflux separator, one or two new pumps (depending on the manner in which a spare pump is provided for each service), instrumentation, piping, and foundation work.

SPLIT-TOWER OPERATION

The basic principle of split-tower distillation is an increase in thermodynamic efficiency resulting from the repeated use of a given quantity of energy. The simplest of many possible applications is illustrated in Fig. 33. Here a series of "n" towers are operated in parallel with approximately equal feed rates, but each at a different (decreasing) pressure. All the required heat is applied to the first (highest pressure) column. Its overhead vapors are condensed by reboiling the next column, and so on. The energy required for the series is about 1/n times the single-column requirement.

There are several constraints, both physical and economic, which limit the number of towers. The highest pressure tower must be below the critical point; its reboiler must not exceed the temperature of the highest available

heat source; and its temperature must not be high enough to cause thermal degradation of the product. The lowest pressure column is usually determined by the pressure at which cooling water can condense the overhead. Between these two extremes, there must be enough differential pressure (and temperature) from column to column to provide sufficient driving force for the condenser—reboiler. In this study only two tower systems were examined.

Energy-Conservation Potential

This option, which was given a detailed examination by Tyreus and Luyben in 1975,²⁷ showed marginal justification using the energy price data available then. The following example is one of their case studies scaled to current energy and equipment prices.

Figure 34(a) shows a conventional methanol—water fractionator for the production of 500 million lb of methanol per year. Part (b) of the figure shows the same separation performed by a two-stage distillation. Note that a 47% reduction in energy was achieved using the split-tower approach, at an incremental capital investment of \$306,000 (scaled to 1978). At current energy prices, that

gives a savings of \$164,000/year and a 1.9-year simple payout.²⁸

The economics for a retrofit situation would be quite different. Note that the original single-stage tower is significantly different from either of the two in the split-tower case. Unless substantial salvage credit is available for the existing tower, the capital investment that must be justified would be very high for a retrofit situation, and the resulting payout period would be long. Such an outlay could probably be justified if it was done concurrent with a planned unit expansion, so that the size of the low-pressure tower would be close to that of the existing column.

Guidelines for Application

The concept of split-tower operation is theoretically applicable to any reboiled distillation column. It can be justified most easily for new units with high reboiler loads. The previously cited article by Tyreus and Luyben identified a breakpoint of about 20,000 lb/hr of reboiler steam usage (which corresponds to about 17 million Btu/hr) above which split-tower operation should be seriously considered. This type of high reboiler load will typically be

Fig. 33 Multi-effect distillation with individual feeds.

encountered when one or more of the following are present:

- High feed rate
- High reflux ratio
- High latent heat of vaporation

Generalized Calculation Procedure

An approximate figure for energy savings for a split-tower system compared with a conventional tower can be made by dividing the reboiler heat duty by the expected number of towers. This provides an order-of-magnitude

estimate. A more detailed figure will require computer simulation of the subject columns.

$$Q_n = \frac{Q_c}{n}$$

where Q_n is the energy consumption for the split-tower system, Q_c is the energy consumption for the single tower, and n is the number of towers.

Published cost correlations are good sources for preliminary cost estimates. Information needed for an estimate includes size of towers, numbers of trays, materials of construction, operating parameters, and specifications on auxiliary equipment (pumps, piping, etc.).

Characteristic	(a)	(b)	
		Column 1	Column 2
Inside diameter, ft	9.5	5.0	7.0
Number of trays	40	60	30
Pressure, psia	17	100	17
Reflux ratio	0.9	1.5	1.01
Reboiler temperature, °F	223	326	223
Reboiler duty, 10 ⁴ Btu/hr	46.7	24.7	0
Reboiler steam, 10 ³ lb/hr	51.3	30.7	0
Steam pressure, psig	50	300	
Cooling water, 10 ³ gal/hr	280	0	152
Utility costs, 10 ³ \$/hr*	990	825.7	
Savings, 10 ³ \$/year		164.3	
Incremental investment, 10 ³ \$		306	
Simple payout, years		1.9	

*Cost of 50 psig steam, \$2.00/10³ lb; 300 psig steam, \$3.00/10³ lb; cooling water \$0.08/10³ gal. Operating time, 330 days/year.

Fig. 34. Economics of split-tower operation for a methanol-water fractionator.

Chapter 6 Other Conservation Options

Distillation is a very broad field, and there are many potential options for energy conservation within it. It is beyond the scope of this document to cover all conceivable options. Emphasis was placed on those with retrofit potential since the older units offer the greatest opportunity for energy savings. Technology that represents a novel approach to distillation energy conservation was also given a detailed review, even though many such options proved difficult to justify economically on a retrofit basis. Other options show great promise, but they are not yet commercially developed.

OTHER AVAILABLE OPTIONS

Several classes of currently available conservation options that were omitted from the detailed discussions are:

- Options that conserve energy but have a long payout period for current and projected energy prices
- Options already in common practice which can normally be justified only at the new design stage
- Alternatives to separation by distillation

Each of these classes and the appropriate options in them are discussed briefly in this section.

Only one conservation option was identified that could not be justified under some reasonable set of conditions—shaft power recovery from waste heat by a thermal (Rankine) cycle engine. This type of waste-heat recovery must compete with the much more efficient process heat exchange. It was assumed that waste heat above the 250°F level generally would be recovered by direct heat exchange. A thermal cycle engine operating at a level below 250°F has a thermal efficiency of only 5 to 10%. For typical waste-heat sources, such recovery efficiency gives a payout in the range of 20 years. Of course, this improves for larger sources because of the economy of scale, but it appears that a waste-heat source well in excess of 100 million Btu/hr would be required to approach a marginal 5-year payout.

Several options were identified which could not be justified for retrofit application and are already in common practice in new distillation design. Among them were:

- Sequencing of towers in a multicomponent distillation train
- Use of side streams in multicomponent distillation
- Use of air coolers in place of cooling water exchangers

Although judicious use of these options can conserve energy, they are already being used where applicable. Consequently, there is little to be gained by exploring them further.

In some cases the best energy-conservation option for a difficult distillation application is the use of a different separation process. Among the candidates for substitution are:

- Extractive distillation
- Azeotropic distillation
- Liquid—liquid extraction
- Adsorption
- Crystallization

Although each of these processes could potentially replace some classical distillation systems, the range of application is very small. A thorough study of these alternatives is far beyond the scope of this text.

EMERGING ENERGY-CONSERVATION TECHNOLOGY

Both government and industry are actively pursuing research in the field of energy conservation. This section is intended as a brief overview of some novel approaches currently under study.

One subject under study is the use of ultrasonic technology in petroleum and chemical processing plants. It has been suggested that ultrasonics could be used for on-stream cleaning of heat exchangers, on-stream decoking of furnaces, atomization of viscous fuel oils, and agglomeration of entrained liquid mist in vapor streams.

The use of microwave heating is also being investigated. It has been hypothesized that microwaves could be used selectively to heat only the reactive components in a mixture, thus saving energy in comparison with bulk thermal heating.

Much study is being concentrated on the use of alternate separation technology in applications where distillation has classically been used. In addition to these alternate separations, such novel approaches as chromatographic separation and separations based on freezing point differences are being evaluated.

Novel approaches to distillation are also being explored. One such project is studying the feasibility of a modified crude distillation process using stagewise vaporization of each successive product.

Another interesting development is the commercialization of the magnetic heat pump. Magnetic heat pumps use solid ferromagnetic materials as "refrigerants" in a Stirling cycle. They have long been used in the laboratory to achieve cryogenic conditions, but only recently develop-

ments have indicated their possible use at near ambient temperatures. Magnetic heat pumps promise higher efficiency and lower cost than heat pumps with conventional liquid-vapor working fluids.²⁸

Chapter 7 Implementation of Energy-Conservation Program

The performance testing, calculations, and different levels of economic options available for distillation energy-conservation programs are described in the preceding chapters. This chapter ties this information together into an overall plan for implementing energy-conservation modifications. It is important that such a study be well organized and efficiently coordinated throughout to keep the cost of the study relatively low. An efficiently coordinated program will also help ensure that proper engineering and economic decisions are made. For example, implementing energy-conservation measures should not hamper existing plant operation and production schedules.

Figure 35 gives the suggested steps for planning and implementing an energy-conservation program. Reducing operating costs in distillation is so complex, with many options and combinations of options available, that planning will help to avoid erroneous conclusions and the necessity to repeat steps.

The first part of the suggested program is a complete performance evaluation as described in Chap. 2. This will provide immediate feedback in possible problem areas in the distillation operation and a wealth of information to be used in further evaluations of the system. Many times changes made in correcting an operating problem can effect enough savings to justify the entire program.

Each distillation system will be different, and special circumstances will arise with each evaluation program. The engineer must use his knowledge of the subject system to adapt the general procedures in this manual.

Associated with each option in this document are guidelines that should quickly give the engineer an idea whether it is applicable to his system. All the options should be screened initially, and then each attractive option should be evaluated individually. Combinations of two or more feasible options should also be evaluated.

A detailed economic analysis for each feasible method should be the last step in the decision-making process. The basis for savings should be accurately calculated, and the purchased and installed costs for any new equipment should be obtained. A vendor or a consulting firm can provide good cost data or the entire economic analysis. As indicated by the dashed lines in Fig. 35, consultants and vendors can play a large part in an evaluation program. Depending on the size and availability of a company's technical staff, contracting for outside assistance is a viable alternative to using limited in-house manpower.

After the economic analysis is complete, a decision is made on what action will follow. The decision will require

different levels of approval depending on the capital investment required. The development of an evaluation program is an excellent mechanism for winning management support for implementing energy-conservation modifications.

Detailed engineering and design can begin with the approval of the project. The project may be designed by the company's own technical staff or by an independent contractor. The information transmitted to the contractor should be very complete, and it is a good idea to meet with him regularly to resolve problems and stay abreast of progress. Once the new equipment is installed and successfully started up, the performance should be periodically checked to ensure optimum operating conditions.

Fig. 35 Steps in implementing an overall energy-conservation program.

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This bibliography represents a compilation of all useful literature found in this study categorized according to options. Although many of these sources were not referenced directly, they provided excellent supplemental material and an enhanced understanding of each option.

It would be advisable for the plant engineer to obtain the sources listed for any conservation option which he plans to carry into a detailed economic and/or technical evaluation.

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