2. RESULTS OF THE STUDY ON THE CHEMICAL INDUSTRY

2. RESULTS OF THE STUDY ON THE CHEMICAL INDUSTRY

2.1 Results of the Study at Tehran Refinery

2.1.1 Outline of the Plant

(1) Plant name

Tehran Refinery Co.

(2) Address

Ghom Road, Tehran

(3) Number of employees

2,140 consisting of

Process engineers:	18
Mechanical engineers:	8
Electrical (or system) engineers:	3
Personnel for maintenance:	120 × 4 = 480
Operators:	350 × 4 = 1,400

Others for management, office work, fire-fighting and security.

(4) Major products

a. LPG (Liquefied Petroleum Gas)
b. Gasoline
c. Kerosene
d. Gas oil
e. Fuel oil
f. Lube oil
g. Asphalt (Blown Asphalt)

(5) Production capacity

The capacity of the atmospheric distillation equipment as the basis for the petroleum refining capacity in an oil refinery, that is, the sum of those of the South Plant and North Plant, is as follows:

a. South Plant: 125,000 BPSD (as designed)
b. North Plant: 100,000 BPSD (as designed)

Note: BPSD = Barrel per stream day

The anual total capacity for processing crude oil is approximately 260,000 BPSD .

(6) Production process

This is an orthodox oil refinery that mainly manufactures the fuel oil and lube oil. Special products such as airplane gasoline and electrical insulation oil are not manufactured.

Configurations of major equipment in the North and South Plants are approximately same but the South Plant only has the lube oil manufacturing equipment. The lube oil fraction taken in the North Plant is transferred to the South Plant. The lube oil is manufactured using the sum of lube oil fractions in both plants.

As a feature of the refining processes, there are the ISOMAX process, PLATFORMER process, VISBREAKER process, and LUBE manufacturing equipment available in addition to the atmospheric distillation equipment and vacuum distillation equipment. All of the licensers for these major equipment are firms in the United States (i.e. UOP, TEXACO, CHEVRON, etc.).

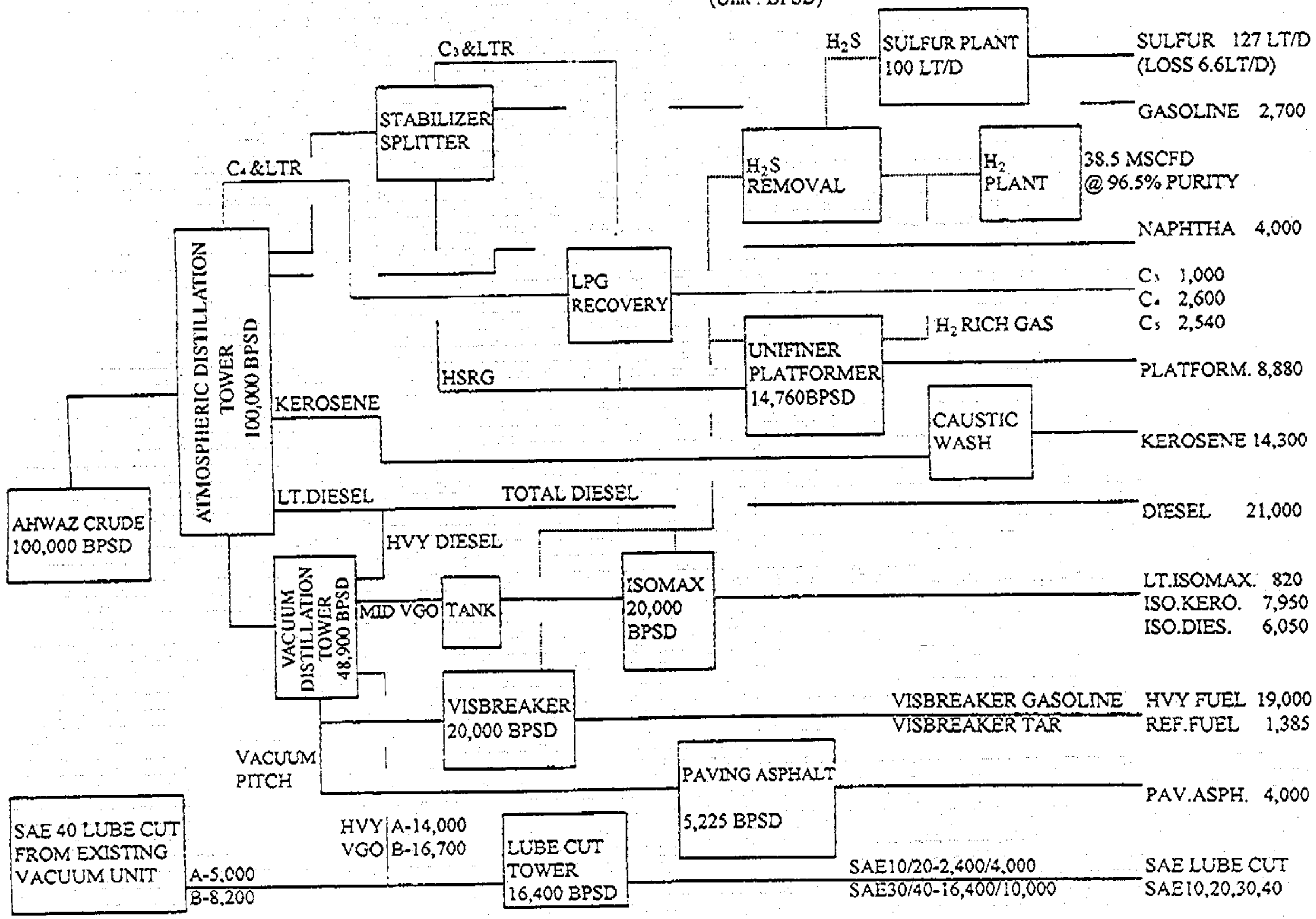
In the course of refining crude oil, high-purity sulfur (S_x; liquid or solid) is manufactured from hydrogen sulphide as a by-product in multiple hydrodesulfurization processes.

Figure 2.1 shows the process block flow in the North Plant.

The fluid catalytic cracking equipment as a typical process for upgrading heavy oil with a catalyst is only provided at Abadan Refinery inside Iran and there are few other heavy oil decomposition processes.

Figure 2.1 Block Flow

(Unit : BPSD)



(7) History of this refinery

Under the Iran's NIOC Plan, the South Plant was constructed 27 years ago in 1968. The North Plant was constructed 21 years ago in 1974. In the meanwhile, major modifications and additional constructions were not made; therefore there are no hi-tech or advanced processing equipment available. All of the existing equipment were built upon initial constructions. Therefore, maintenance has become important for stable operation.

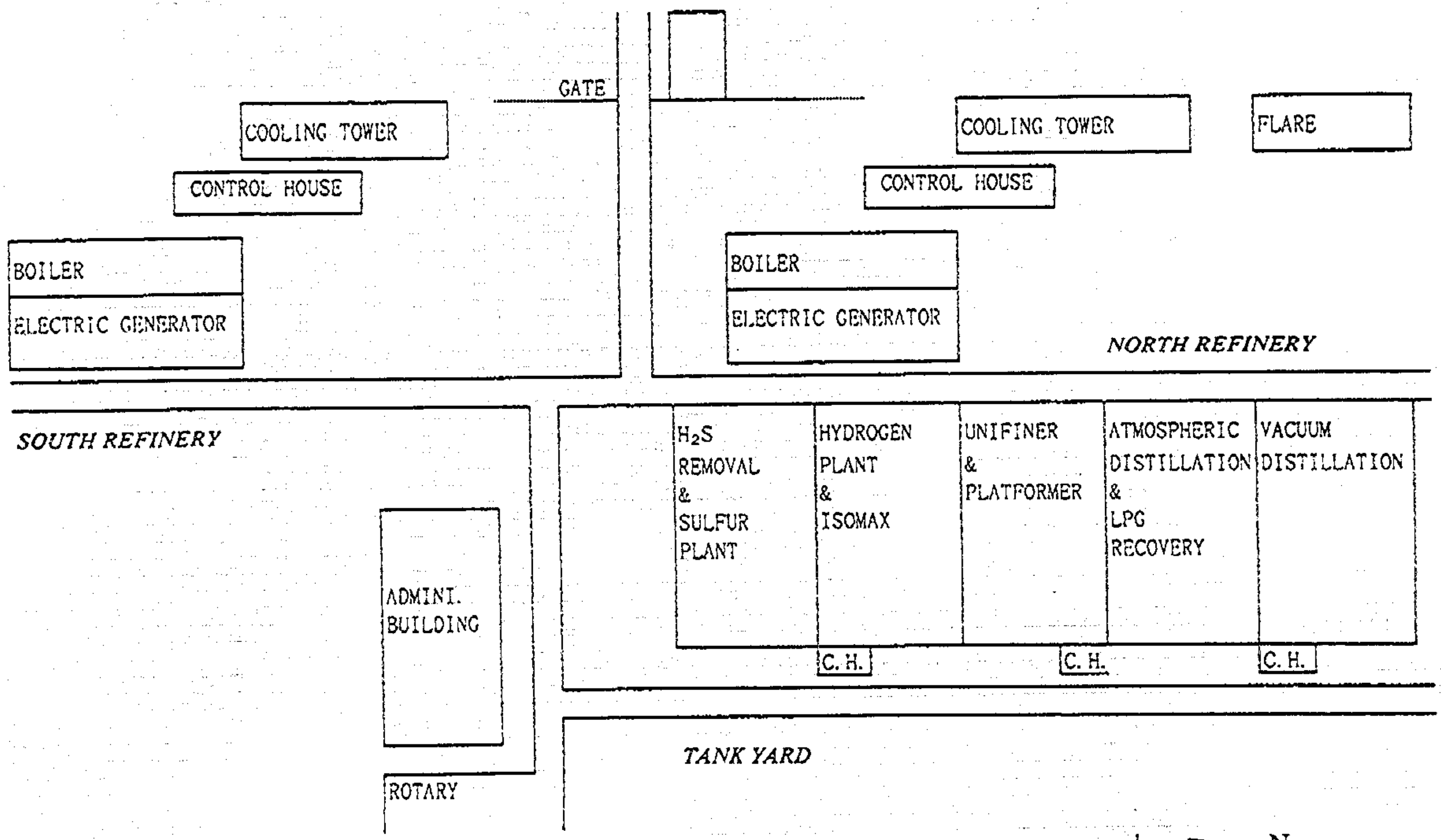
As described earlier, the nominal capacity of both plants for processing crude oil is totally 260,000 BPSD. This refinery is the third largest one in I.R. Iran and is an important one targeted for the markets in north cities in I.R. Iran along with Arak and Tabriz.

(8) Plant layout

Figure 2.2 shows the layout of the entire north plant that we have investigated. The space between equipment is several times larger than that in Japan, and the entire refinery has been constructed in a sufficiently wide area.

In the South Plant, the lube oil manufacturing equipment is provided in addition to those in the North Plant. Flare stacks of both plants are symmetrically laid out, and the equipment in both plants are at approximately symmetrical locations.

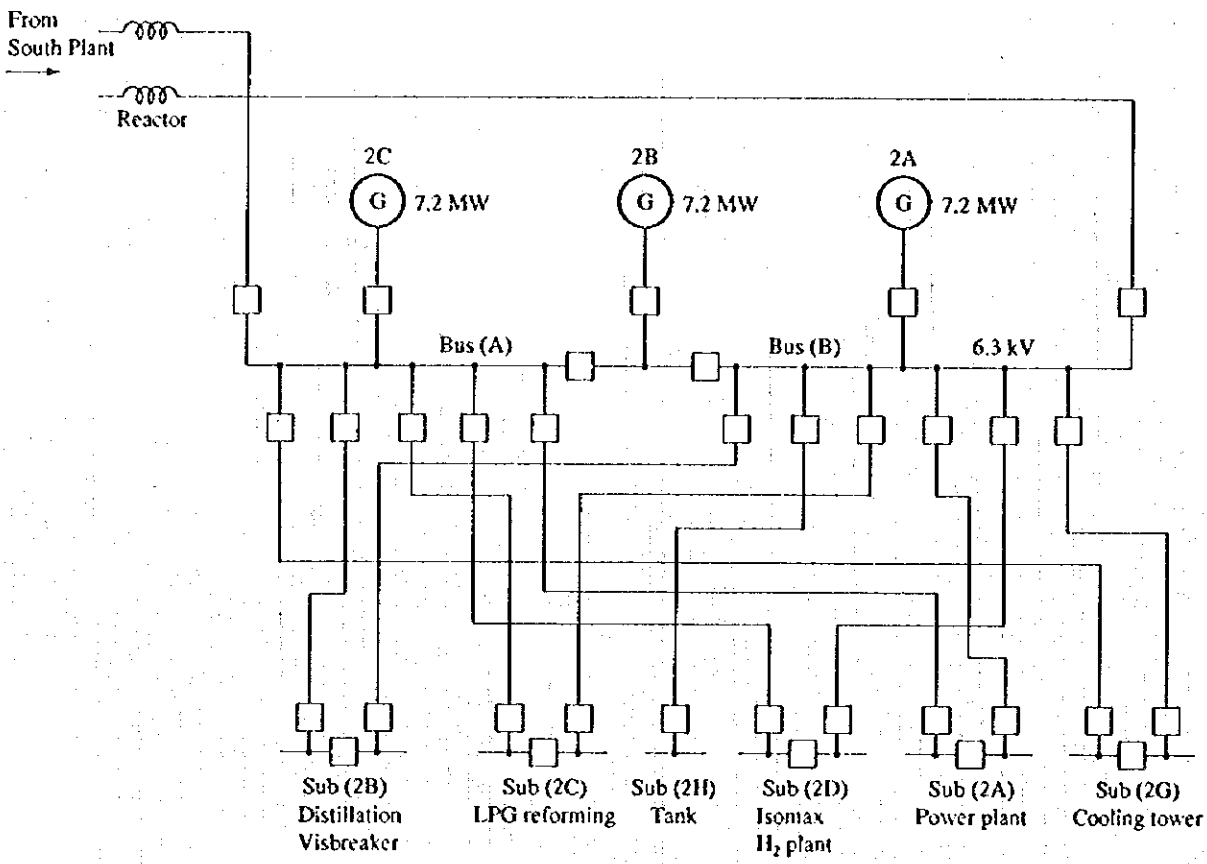
Figure 2.2 Plant Layout



(9) One line diagram

Figure 2.3 shows one line diagram.

Figure 2.3 One Line Diagram



(10) Description of major equipment

For major equipment, the following equipment are provided in addition to the atmospheric distillation equipment and vacuum distillation equipment as the basis for the oil refinery:

a. Atmospheric distillation equipment and vacuum distillation equipment

The atmospheric distillation equipment separates the rough-distilled product from crude oil through distillation at a tower overhead pressure of approximately 1.5 kg/cm² (G). This refinery uses an ordinary pipe still continuous distillation process consisting of a heating furnace and a main rectifying column.

The vacuum distillation equipment processes the fraction from the bottom of the atmospheric distillation equipment. Since, if crude oil is heated to a temperature higher than required in atmospheric distillation, oil quality deterioration is caused by thermal decomposition, this equipment distills under vacuum (approximately 30 mmHg abs) so that high boiling-point component will be evaporated in a relatively lower temperature. This refinery adopts an ordinary vacuum distillation process using a steam ejector.

b. LPG fraction recovery equipment

This equipment recovers and refines propane (C_3), butane (C_4), pentane (C_5), etc. light hydrocarbon fraction from the atmospheric distillation equipment and straight-run naphtha stabilizer.

c. Catalytic reforming equipment (UNIFIER & PLATFORMER)

This equipment consists of two processes; the UNIFIER section that desulfurizes the low-grade gasoline (straight naphtha) fraction to be fed and the PLATFORMER section that manufactures a high-RON (Research Octane Number) gasoline blend stock by applying reforming effects (aromatization and hydrogenation/decomposition) on desulfurized gasoline.

Since activity of the reformer catalyst reduces day by day, operation is stopped and the PLATFORMER catalyst is regenerated after a certain period. The licenser for this process is UOP.

d. Hydrogenation desulfurizing equipment (ISOMAX)

This equipment mixes the raw material oil of kerosene and gas oil fractions with hydrogen gas and causes denitrification and deoxygenation as well as hydrogenation and desulfurization as a result of reaction at a high temperature and a high pressure under existence of a catalyst.

The licensers are UOP and CHEVRON in the United States. Heavy metals such as vanadium (V) and nickel (Ni) in the raw material oil will generally form catalyst poisons. The one-stage reactor is used for reaction with the amorphous catalyst having the metal resistant capability against these poisons.

e. Visbreaker

This is a viscosity reduction equipment that applies heat treatment to the fed high-viscosity fuel oil under relatively mild operating conditions to obtain low-viscosity oil by breaking paraffin chains of naphthene and aromatic compounds.

f. Sulfur recovery equipment

This equipment consists of the amine absorbent recirculation unit that recovers hydrogen sulfide (H_2S) generated in desulfurization processes of the UNIFIER and ISOMAX described above and the sulfur recovery unit (i.e. sulfur unit) that converts the recovered H_2S into liquid sulfur (S_x) by the Claus reaction.

The designed capacity of the sulfur recovery equipment is 80 tons/day (sulfur recovery rate 95 %) but the actually recovered volume is about 40 tons/day (according to sulfur content in the crude oil processed).

The remaining sulfur in the tail gas of this equipment is burned by two incinerator using the by-product gas from the asphalt equipment, and converted into SO_x (normally approx. 4 t/d) to be released into the atmosphere.

g. Lube oil manufacturing equipment

The South Plant has the following lube oil manufacturing equipment. The licenser for these lube oil equipment is TEXACO in the United States. Twenty-three years have passed since these equipment were constructed.

1) Furfural extraction unit

By using the excellent selective resolution capability of the furfurylaldehyde (C_4H_3OCHO) solvent at 80 °C to 120 °C, this unit removes aromatic compounds and naphthene compounds from the raw lube oil to extract high-grade lube oil component.

2) MEK dewaxing unit

Wax in raw oil reduces lubrication performance at a low temperature. Therefore, by using MEK (methyl-ethyl-ketone; $CH_3COC_2H_5$), the raw material oil is resolved, cooled, and then the wax precipitated is removed by filtering for refining.

3) Hydrofinishing unit

Hydrofinish is performed to improve the color, odor, and stability against oxidization of the lube oil. Since the operating conditions for this method are milder than the ordinary hydrofinishing method for fuel oil, the running cost is lower and the yield is higher.

h. Boiler

The North Plant has three boilers, while South Plant has four boilers. The total volume of generated steam is 1,600,000 lb/h (720,000 kg/h), among which 300,000 lb/h (135,000 kg/h) is used for power generation and the rest is consumed in processes. The fuel used is the fuel oil produced in this refinery and the fuel gas.

In the South Plant, three boilers and a power generator are normally running (i.e. one of four boilers is out of service for overhauling) The steam piping in the South Plant is connected to that in the North Plant for flexible use of steam.

Due to the limitations of the heating pipe strength against pressure and of the water processing equipment, three boilers are normally running at a 70 % load.

i. Electric equipment

Electric power used by this refinery is normally fully supplied from the private power generating plant. The power line in the refinery is usually not connected with an external power line having problems in reliability. Only in emergency cases such as power failure of the private power generation plant, power is purchased from the external power line. Purchased power is supplied at 20 kV to four 500 kVA power receiving transformers. The total capacity is only 2,000 kVA, so it is impossible to supply all power for the refinery. This refinery is divided into the North Plant and South Plant, and the power lines are almost symmetrical. Each power line has three 7.2 MW power generators. Among these six power generators, five generators are always running and the rest is provided as a standby generator. The North and South Plants are connected with each other by using two 6.3 kV lines, each of which passes through a 3,750 kVA reactor. The bus voltage of the power plant is 6.3 kV as the voltage of the power generator. Power is always supplied from both bus lines A and B to each plant (excluding the 2H substation plant) via two lines. No synchronous motor and power capacitor are installed, so the power factor improving actions have not been taken.

(11) Energy price

Both fuel and power are produced in the refinery. Nothing is purchased.

(12) Study period

June 15 through 19, 1996

(13) Members of the study team

a. JICA team

Leader	: Norio Fukushima
Process management technology	: Kenji Kazuma
Heat management technology	: Jiro Konishi
Heat management technology	: Seiichiro Maruyama
Electricity management technology	: Kazuo Usui
Economic evaluation	: Shigeaki Kato (Preliminary study)

b. PBO team

Energy conservation	: Mr. A. Mazhari
Energy conservation	: Mr. S, Akhavan
Macro level energy management	: Mr. Azizi
Macro level energy management	: Ms. Zarvani
Macro level energy management	: Mr. Mohamadzadeh (Preliminary study)

(14) Interviewees

Mr. Mikailian	Head, Operational Engineering Department
Mr. Mahtaji	Head, Process Enegineering Department
Mr. Fathi	Head, Distillation Department (Preliminary study)
Mr. Zareh Dashti	Head, Power Utility Department
Mr. Ebrahimzadeh	Process Engineer
Mr. Tajik	Process Engineer
Mr. Tehrani	Process Engineer
Mr. Alavi	
Mr. Aref Dowlatabadian	NIOC Refineries Expansion Section
Mr. Faridi	Manager, Material Department (Preliminary study)

2.1.2 Energy consumption status

(1) Production

Table 2.1 shows the total crude oil volume processed and the major products produced at both North and South Plants in 1989 through 1994. Since this refinery has not made equipment modification, the crude oil volume processed in this period was the high level and average operation rate between 91 and 94 %, and the quality of the processed crude oil was almost not changed. Therefore, the throughput of each product remained almost at the same level.

Table 2.1 Production of Major Products

Name of Product	1989 Amount 1,000 m³	1990 Amount 1,000 m³	1991 Amount 1,000 m³	1992 Amount 1,000 m³	1993 Amount 1,000 m³	1994 Amount 1,000 m³	1994 Yield %
CRUDE	13,671	14,126	13,776	13,724	13,572	13,981	[97.9]
LPG	417	438	404	381	440	432	3.1
Gasoline	2,008	2,002	2,021	1,972	1,998	2,461	17.6
Kerosene	2,008	2,151	1,802	1,833	1,789	1,893	13.5
Gas oil	3,441	3,206	3,380	3,298	3,402	3,500	25.0
Fuel oil	4,074	4,345	4,372	4,789	4,456	4,410	31.5
Lube cut	382	382	456	375	420	386	2.8
Asphalt	579	671	605	375	676	611	4.4

(2) Service Factor

Since each equipment is out of service for approximately one month in every three years for normal maintenance, the average service factor is apparently:

$1 - (1/36) = 0.972$

This rate is higher than that in Japan, where each equipment is out of service for approximately one month in every two years (i.e. apparent operation rate: $1 - (1/24) = 0.958$). However, the real average rate presumed from Table 2.1 is between 0.91 to 0.94, which is lower than the level in Japan ($\fallingdotseq 0.95$). The reasons seem to be effects of emergency shutdown due to operation troubles.

(3) Energy consumption

Table 2.2 shows the utilities consumed by both plants.

Table 2.2 Annual Utilities Consumption

Name of Utilities		Consumption					
		1989	1990	1991	1992	1993	1994
Natural gas	(10^6 m³)	417	315	428	466	335	578
Fuel gas	(10^6 m³)	456	688	553	378	372	333
Fuel oil	(1,000 m³)	346	336	234	295	272	319
Slop oil	(1,000 m³)	172	173	149	106	78	122
Electricity	(MWh)	200,527	177,780	189,403	196,456	182,220	188,650
Water	(1,000 m³)	7,426	7,465	7,751	7,716	6,784	6,570

Generally, the refinery uses non-condensing off gases (H_2, C_1, C_2, etc.) discharged from each equipment as fuel. Therefore, consumption of natural gas, etc. introduced from the outside is small.

There is no other plant in the surrounding area of this refinery. Therefore, they can not enjoy the opportunity of improvement of energy efficiency by such means as passing raw materials to or from surrounding plants through the pipeline or sharing various utilities, as seen in industrial complexes in Japan.

(4) Energy consumption by process

The meters of utilities are installed at the supply side of each utility but no meter is provided at each consuming site. Therefore, we could obtain only partial measurement data in this survey. Further, each utility is balanced between the North and South Plants and some utilities move between both plants; therefore it is impossible to catch the real consumption at each plant. Since actual running data could not be obtained, Table 2.3 shows the designed energy consumption by equipment in the North Plant.

Table 2.3 Energy Consumption

Base: Design value

Unit	High Pressure #650 Steam t/h	Med. Pressure #300 Steam t/h	Low Pressure #60 Steam t/h	Electricity kW
Atmospheric & *1 vacuum distillation & lube	11.5	15.7	14.8	3,163
LPG unit	–	9.4	6.8	452.2
Isomax	62.7	3.0	(+25.4)*2	1,197.5
Platformer	14.7	0.1	0.01	773
Visbreaker	–	5.6	0.3	1,164
Roofing asphalt	–	0.9	–	–
Paving asphalt	–	10.7	–	16.6

Notes *1: Atmospheric distillation (100,000 BPSD)
*2: Steam production by extraction
: psig

Table 2.4 shows the power supply/demand status of the North Plant. At the point of survey, power generator 2A was out of service and two power generators, 2B and 2C, were running. For the load-side power, the numeric value shown on the cumulative power meter on the power distributing panel is used. For the power factor, the value measured at a different point of time is used. Power consumed by the entire North Plant is approximately 12 MW and the power factor is 82 %.

Table 2.4 Power Supply and Demand Status (North Plant)

15:00 Jun. 17 to 8:00 Jun. 18

(Supply side)

Equipment Name	Active Power (MW)	Reactive Power (Mvar)	Power Factor	Apparent Power (MVA)	Remarks
Generator 2B	4.98	3.04	0.853	5.83	
Generator 2C	4.97	3.94	0.783	6.34	
Generator total	9.95	6.98	0.819	12.15	
South plant Delivery total	0.06	0.12	0.46	0.13	Demand and supply balance adjustment
Supply total	9.89	6.86	0.82	12.04	

(Load side)

Equipment Name	Active Power (MW)	Reactive Power (Mvar)	Power Factor	Apparent Power (MVA)	Remarks
SUB2A(A)	0.83	0.46	0.87	0.95	
SUB2A(B)	0.63	0.44	0.81	0.77	
SUB2A total	1.46	0.90	0.85	1.72	
SUB2B(A)	2.25	1.51	0.83	2.71	
SUB2B(B)	1.54	1.20	0.79	1.95	
SUB2B total	3.79	2.71	0.81	4.66	
SUB2C(A)	0.88	0.59	0.84	1.06	
SUB2C(B)	0.68	0.52	0.79	0.86	
SUB2C total	1.56	1.11	0.82	1.91	
SUB2D(A)	0.90	0.61	0.82	1.09	
SUB2D(B)	0.69	0.35	0.76	0.91	
SUB2D total	1.59	0.96	0.85	1.86	
SUB2G(A)	0.65	0.49	0.8	0.82	
SUB2G(B)	0.36	0.27	0.8	0.45	
SUB2G total	1.01	0.76	0.8	1.26	
SUB2H	0.48	0.42	0.75	0.64	
Load total	9.89	6.86	0.82	12.04	

Since power supply depends on private power generation only, safe and stable operation with the refinery is important when energy conservation is considered. At the point of study, the equipment utilization factor of the power generators (= average power/ generator rating output) was 69 %, which means a significant allowance. For the entire refinery, it may be possible to decrease the number of power generators from 5 to 4. However, three generators cannot supply the entire load and there is a limitation on purchasing electric power. Therefore, a problem may arise when a generator is out of service. When the bottle-neck in supplying parts is considered, the light-load operation may be inevitable to reduce the potential of troubles.

Table 2.5 shows the load fluctuation in each substation at the point of study. Substations excluding 2H are in a stable load status.

Table 2.5 Load Fluctuation at Each Substation

Substation Name	2A	2B	2C	2D	2G	2H	Total Load
Maximum power (MW)	1.56	3.80	1.57	1.62	1.02	0.25	12.13
Mean Power (MW)	1.53	3.75	1.45	1.62	1.01	0.17	11.74
Minimum Power (MW)	1.48	3.68	1.38	1.61	1.01	0.04	11.13

(5) Energy intensity

At the site using each utility, there is no meter provided. Energy consumption on a practical basis can only be checked for the entire refinery.

Level of energy intensity is as follows:

Electricity:	14.4 kWh/m³ - crude oil
Steam:	431 kg/m³ - crude oil
Industrial water:	532 L/m³ - crude oil
Compressed air:	9.37 m³/m³ - crude oil

Since real operation data other than a part of the heating furnace capability and the energy consumption of the entire refinery could not be obtained, it is not possible to compare the energy intensity with that of plants in Japan.

(6) Energy flow

Figure 2.4 shows the energy flow of the entire refinery.

Figure 2.4 Energy Flow

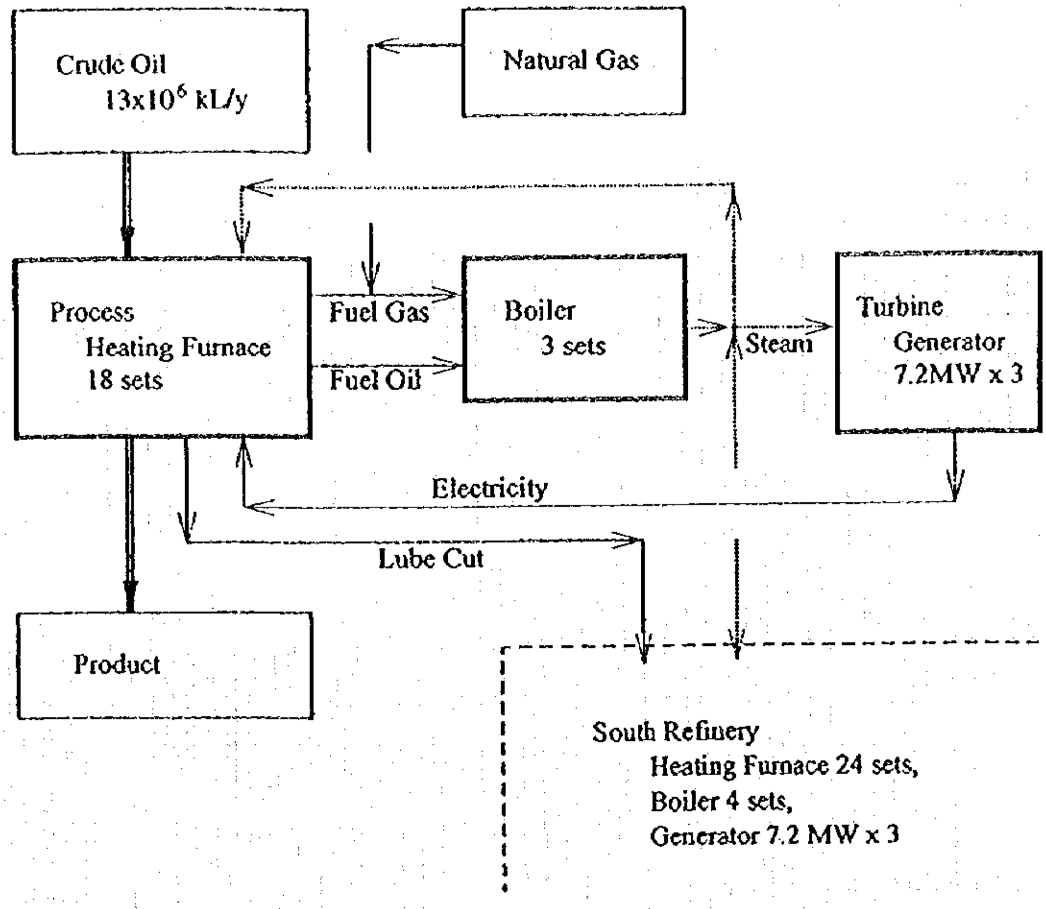
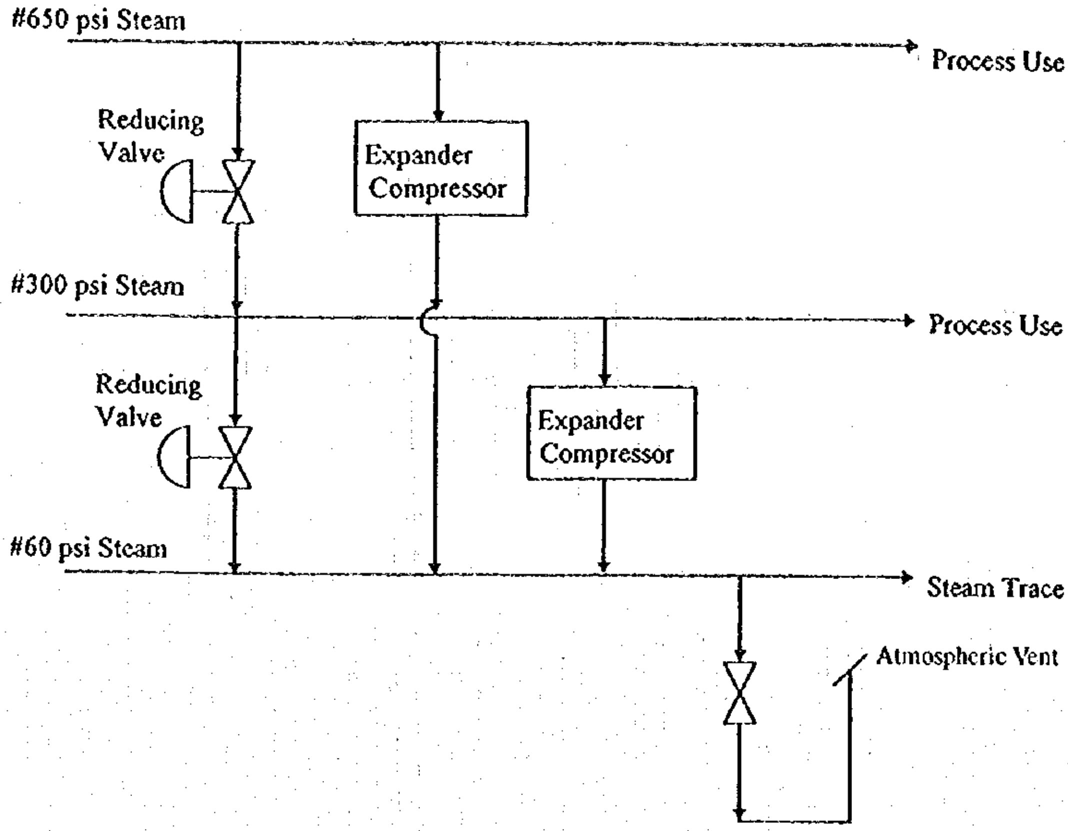


Figure 2.5 shows the steam system in this refinery.

All compressors in this refinery are driven by steam. In summer season, low-pressure (60 psig) steam is purged into air.

Figure 2.5 Steam System



2.1.3 Energy Management Status

To implement operation for energy conservation, it is important that each employee throughout the entire plant finds and thinks about problems and take proper actions for improvement in an organized manner.

For this purpose, the promotion system and each employee's consciousness on energy conservation are vitally important. It may be difficult to achieve effective energy conservation unless the grass-roots' energy conservation improvement activities through the target management, proposal system, etc. is settled as much as possible.

(1) Setting the energy conservation target

In I.R. Iran which is a major oil producing country, the energy (petroleum) price has been low and energy conservation in this refinery has not been considered since it was constructed. Therefore, a meter is provided at the process fluid side to control produced amount but meters are almost not provided for each utility.

To promote energy conservation, it is essential to grasp the real data of energy consumption through inspection. For the current status of this refinery, components and systems for controlling the energy use in detail are hardly provided. For a while, it is necessary to set the control target value based on existing components and promote energy conservation on a software basis by preparation of operation manuals, etc. According to the plant people, approximately 100 new meters for oil and approximately 50 meters for gas are required for energy conservation. And they said the total investment cost amounts to approximately US$ 450,000, which means it can hardly be realized.

In this regard, the Japanese team explained that each item, where installation of metering equipment seems necessary, is evaluated based on the investment cost and profit, and higher priority items (higher merits and shorter payback period) within the scope of budget are selected and implemented sequentially and that with this method, the profit may be large even though the investment is low.

(2) Organized activities

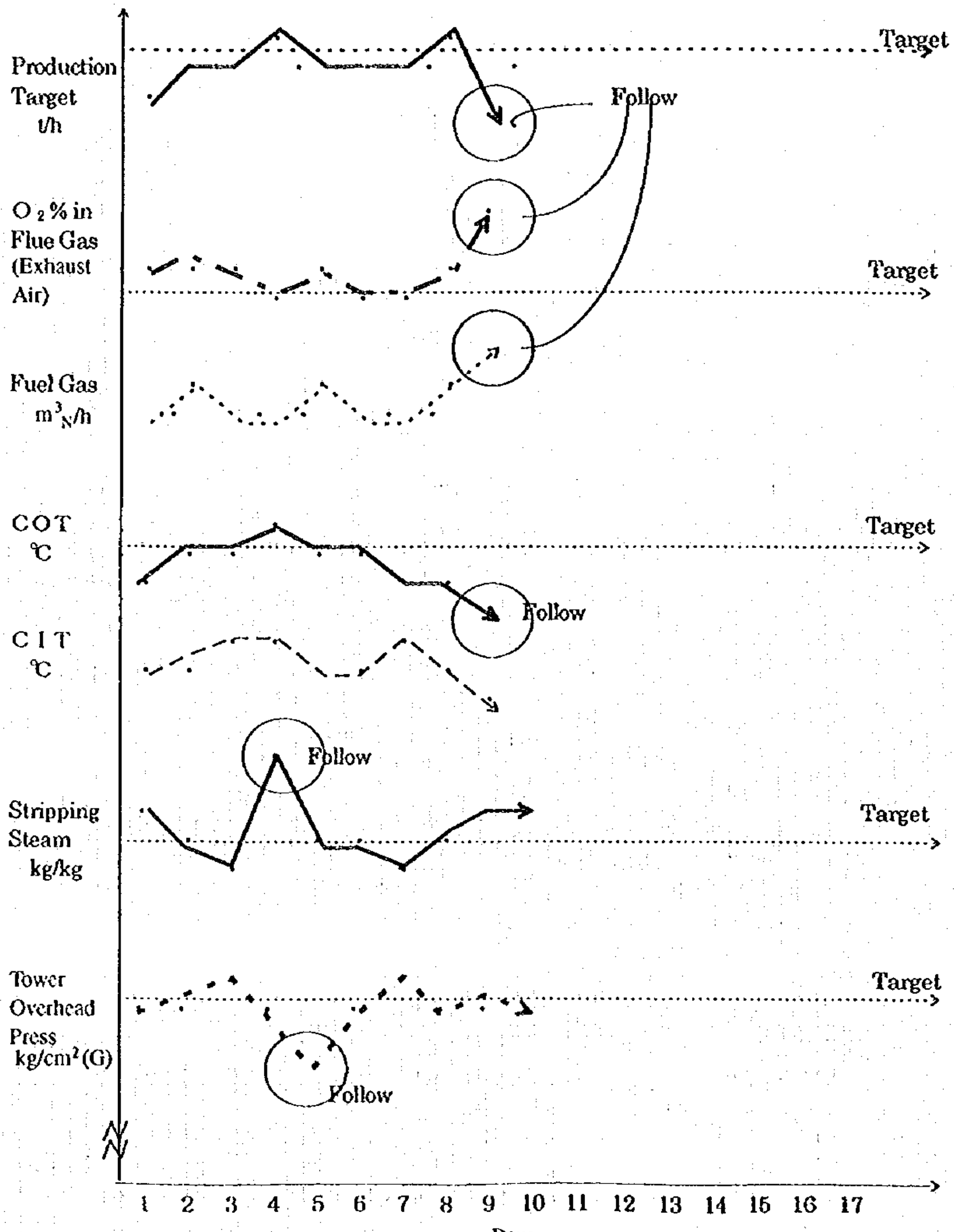
In this refinery, the Energy Management Committee was organized after the point of preliminary study in October, 1995, but actual activities have not been performed since then. In this study, we found that, as the only activity of this committee, the steam leak prevention team was established and leaking locations in the South Plant were being listed. The result is expected.

To promote energy conservation in a systematic manner, collaboration between this committee and related departments (i.e. administration department, engineering department, operation department, etc.) is required. For listing the steam leaking locations, as much cooperation of related group as possible is necessary.

For promotion of energy conservation in Japan, the proposal system plays a big role. Practically, the proposal system is supported by the target management system and aims at troubleshooting of problems that arise within the energy conservation activities. Figure 2.6 shows an example of the operation target management chart that is actually used in a refinery.

For target management, operation control targets relating to process operation variables (flow rate, temperature, pressure) are set, a manager is designated, and then a graph is plotted every day. If a failure (i.e. target overflow) occurs in operation data, the data is immediately followed up, and the action to be taken is examined. Actions that can be immediately taken should be implemented immediately. Actions that cannot be taken immediately but require equipment modification, etc. should be examined by as many as people concerned to summarize a proposal. Table 2.6 shows an example of the proposal form.

Figure 2.6 Operation Target Management Chart



COT : Coil Outlet Temperature
CIT : Coil Inlet Temperature

Table 2.6 Improvement Proposal Form

Date ________________,

Title	Dept.:
	Name:

Current problem: (Describe quantitatively and specifically the matter to be improved. Please attach the actual data, if available.)
Improvement proposal: (Describe your improvement proposal quantitatively and specifically. Please give data analysis results, if available.)
General comment: (Describe the comment of the responsible reviewer.)

Evaluation items:		Evaluation result:
Economical efficiency	¥/Year	Grade ________
Workability improvement	Person/Year	(Grade 1, Grade 2, Grade 3, Grade 4, Grade 5)
	(¥/Year)	

The proposal submitted is evaluated by the related engineering group, etc. and reviewed by the specific organization. The proposal adopted may be awarded or reflected to personnel evaluation to stimulate the people's participation for further proposals.

To promote energy conservation, all employees or as many as employees must participate in energy conservation activities. As many as people should find as many as problems from each standpoint and make comments or submit improvement proposals. Incorporating target management will result in detection of problems and proposals for them directly associated with daily work.

It is essential that the related administration group, engineering (maintenance) group, and operation group cooperate with each other and that each responsible person recognizes himself/herself as a professional on his/her jobs to find potential problems and improvement points.

(3) Control based on data

To eliminate useless energy consumption by reducing process steams (stripping steam and injection steam), etc. and controlling the distillation tower reflux ratio and over-flash volume as important energy conservation items in the refinery, control must be enhanced by cheking the real data on existing equipment.

For this purpose, maintenance of existing metering components should be performed periodically to maintain each accuracy, and inspection on each equipment should be conducted to review process control standard (temperature, pressure, flow rate, etc.) of important control point and revise them accordingly. For the reviewed values, the operation manual and standard should be modified and the values should be reflected to control using the operation check sheet described below. In this plant, the flow meter is not provided at the site where each utility is used; therefore the value indicated by the related measuring component or the opening degree of the control valve should be used as a substitute for control.

When the operating condition is changed, energy conservation can be accomplished by selecting the optimum operation pattern matching that revised operating condition as soon as possible. Regular maintenance and review of operation manuals are important to accelerate it.

After the process control values are reviewed, it is necessary to correct the reviewed items in the operation check sheet and check to see if the new control values are achieved every day after patrol. If any control value is not achieved, the necessary action should be immediately taken. When the immediate action is impossible, the plan to treat with should be clarified at first and proper action is taken later at a proper timing.

Operation control using the check sheet shown in Table 2.7 allows also detection of problems directly associated with daily equipment operation, and items to be improved for energy conservation can be found.

Table 2.7 Check List of Daily Operation

Tag No.	Service	Operation Target	Operation Data	Follow Yes/No	Remark
Pic-AAAA	Tower top press.	*** psig	*****		
TC-BBBB	Reflux drum	*** Deg.F	*****		
Frc-XXXX	Side product	*** BPSD	*****		Flow meter
Fi-YYYY	Stripping steam	*** lb/hr	*****		Flow meter
C-ZZZZ	Stripping steam rate	*** lb-STM/lb-Prod.	*****		Calculated value

With the side stripper, flow meters for the stripping steam and each product were both installed for controlling the initial boiling point of the product since the time of construction. Therefore, management (energy conservation) control is implemented for this item.

Management based on measured data is important for energy conservation. The daily operation check list (Table 2.7) should include energy conservation control items and be combined with chart control (Figure 2.6) so that quantitative management can be performed as much as possible. We expect that such manner of control will result in improvement proposals (Table 2.6).

(4) Education and training of employees

In this refinery, almost no energy conservation activities have been implemented, nor has employee training been conducted. To promote energy conservation, it is necessary to create a system in which each employee think about improvement. For this purpose, raising of each employee's consciousness is finally required.

Employee training on energy conservation will be an important issue in the foreseeable future.

The number of employees is about twice larger than that in similar plants in Japan. However, the number of engineers in each process, mechanical, or electric field is a half or less than that in Japan. Instead, employees engaged in maintenance and operation are several times larger. In other words, this refinery has personnel intended for equipment maintenance to maintain the stable production of petroleum products.

Therefore, the current engineer organization has a limitation when this refinery originally implements various technical studies for efficient improvement on energy etc. to put hardware improvement into practice, in addition to improvement of the operation efficiency using software. If development of original energy conservation technology may be desired for this refinery in future as implemented in plants in Japan, it will be necessary to improve personnel organization by training engineers on a long-term basis.

(5) Equipment management

This refinery has extremely many industrial water leak and steam leak locations compared with Japan. Also, the heat insulating material is broken, and part of the steam tracer is exposed or leak from the gland seal for the pump or valve was found.

When we were measuring temperature distribution around the cooler in the field, leak of a significant amount of kerosene vapour from the body flange of the kerosene reboiler was found. Although the kerosene vapour was slightly grayed and different from pure white steam, it may have been difficult to find the kerosene vapour leakage in the environment where steam leaks exist every day.

According to our notice, the kerosene vapour was reported to the maintenance group. (The level of this leakage requires emergency stop and maintenance of the equipment in Japan.)

As described above, the equipment control status of this refinery is apparently poor. It is necessary to improve the equipment control system in the future energy conservation activities.

2.1.4 Problems with Energy Use and Countermeasures

(1) Comparison with a new Japanese plant

a. Complexity factor and energy efficiency

Since detailed data on the throughput and energy consumption for each equipment was not obtained, no comparison with a similar Japanese oil refinery could be conducted in detail.

As a method of evaluating the energy efficiency of an oil refinery, a complexity factor complementing for the equipment constitution is used.

For your reference, a Nelson complexity factor was estimated by dividing the amount treated by each equipment in proportion to the design value based on the actual crude oil treated amount. As the result of calculation, an approximate value 6.734 was obtained as a complexity factor for the North Plant as shown in Table 2.8.

Table 2.8 Complexity Factor

Name of Unit	Complexity Factor*	Operation	Complexity
		(Feed kL)	
Atmospheric crude distillation	1.00	15,898	1.000
Vacuum crude distillation	2.00	7,774	0.978
Visbreaker	2.00	3,180	0.400
Hydrocracking (Isomax)	6.00	3,180	1.200
Catalytic reformer	5.00	2,347	0.738
Hydrogen production	3.00	6,121	1.155
Asphalt (Blown)	2.00	906	0.114
LPG recovery	1.50	976	0.092
Stabilizer splitter	3.00	2,347	0.221
Caustic wash	2.00	2,273	0.214
Lube cut tower	2.00	2,605	0.328
Total			6.734 (CF)r

*: Revised complexity factor

The approximate value η for the entire oil refinery was calculated by using the energy consumptiom for the entire oil refinery on the assumption that the both South and North Plants have nearly the same complexity factor from their similarity in the facility constitution.

$$\eta = Fr/(A \times (CF)r)$$

where

Fr: Total fuel volume consumed by the equipment in the refinery
Fr = (Total fuel volume; kcal)/9,250 (kcal/L)
A: Volume processed by the atmospheric distillation equipment (kL)
(CF)r: Complexity factor of the refinery

Here,

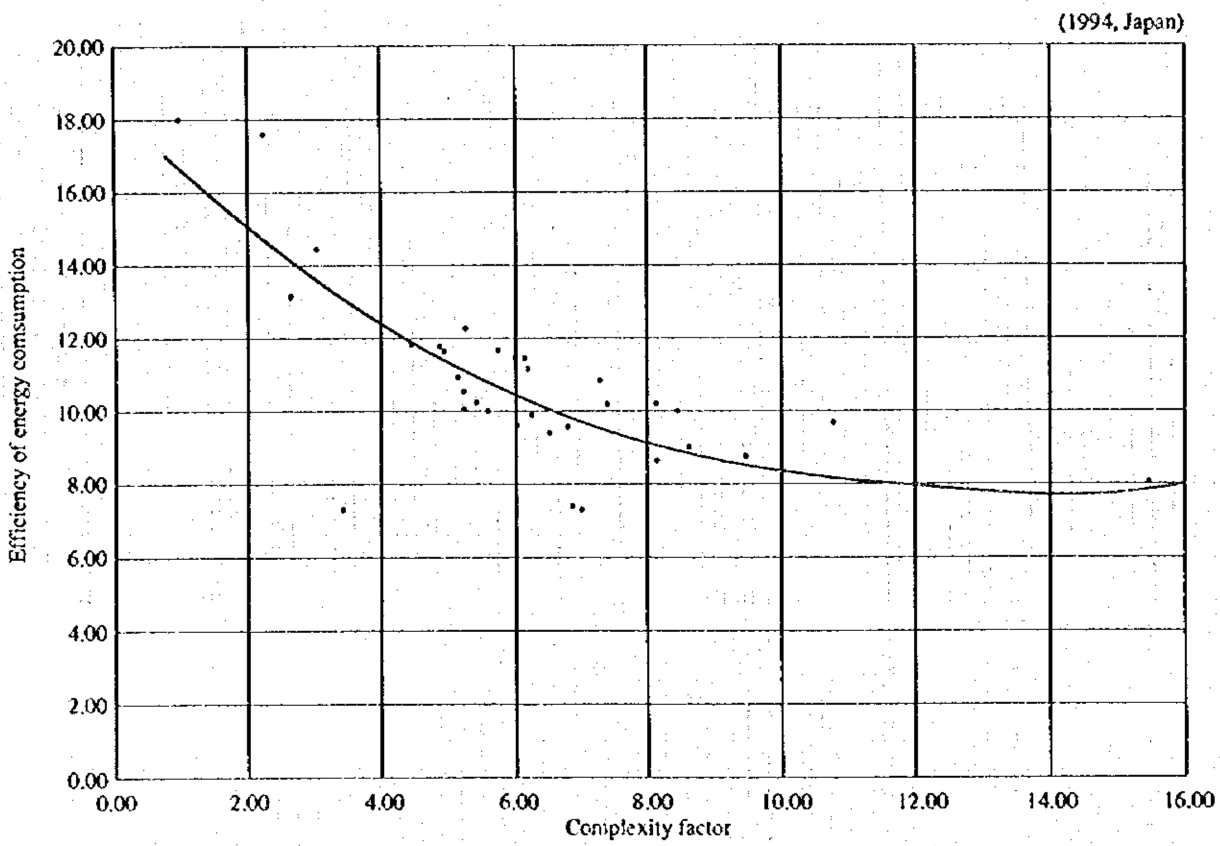
Fr = 1,433 × 10^6 L (based on the actual consumption in 1994 shown in Table 2.2)

Therefore,

$$\eta = (1{,}433 \times 10^6) / [(13{,}981 \times 10^3) \times (6.734)]$$
$$\approx 15.2$$

As a result of the calculation, η of both North and South Plants is approximately 15.2. In the correlation diagram of η and production volume as shown in Figure 2.7, it is about 1.5 times larger than the example (average $\eta \fallingdotseq$ about 10.0) of the equivalent-scale refinery in Japan (15,900 kL/d in terms of crude oil).

Figure 2.7 Complexity Factor vs. Efficiency of Energy Consumption



(Source: Petroleum Association of Japan)

b. General situation of energy use

During normal operation in this refinery, the tank for feeding crude oil to the atmospheric distillation equipment is switched within five to seven days but the crude oil type (mainly Iranian Heavy) and its properties are hardly changed, and variation in the delivered oil volume is small. Therefore, compared with oil refineries in Japan that import crude oil from many countries and adjust operation according to a variety of production needs, energy control is easy in this refinery. However, due to the inexpensive energy cost in Iran, necessity of energy conservation has not been recognized, and actions for energy conservation on the equipment have not been taken into account since the construction of the refinery (e.g. a meter not installed at each equipment site using the utility).

Also, equipment modification for energy conservation has not been implemented Compared with plants in Japan, there are many steam leak points in this plant (See (6)). For the current status of this refinery, the daily energy management system may be said to be inadequate.

To improve the total energy consumption efficiency in this refinery, efforts to increase the volume of oil delivered should be made by taking the debottlenecking actions such as improving the cooling capability of the cooler on the tower top of the vacuum distillation equipment (See (5).) and so on from the viewpoint of hardware on the entire equipment of this refinery. Also, meters should be installed for utilities and various energy modification should be implemented. From the viewpoint of software, efficiency should be maintained by enhancing operation control.

c. Key points in energy conservation

In Japan, energy conservation actions are generally divided into the following three steps depending on the required investment:

Step I: Management enhancement
Step II: Equipment addition
Step III: Process changes

In the manufacturing sites, these action items from step I to III are implemented step by step.

Generally, the process fluid is heated up to a specified temperature in oil refining. In this case, "required combustion" in the heating furnace is represented by the following formula:

(Required combustion heat) = (Required heat of the process fluid) + (Heat loss)
− (Heat given to the process fluid by preheating)

where "Required heat of the process fluid" means the calories required for processes such as distillation and reaction of raw material oil in the processes after the furnace. It is a fixed value required for the processes.

This total amount is unchanged unless the processes are drastically changed even though energy conservation is promoted.

Hence, to reduce the "required combustion heat", "heat loss" should be reduced or "heat given to the process fluid by preheating" should be increased. "Heat loss" includes not only heat radiation from the furnace wall but also exhaust gas loss. "Hert given to the process fluid by preheating" means the calories transferred to the raw material oil fed to the heating furnace. If the feed oil is preheated, "Required combustion heat" in the heating furnace can be reduced by the preheating value.

For energy conservation in oil refining, activities from both aspects, software and hardware, are important, that is, operation control in terms of software and equipment improvement in terms of hardware. With these activities, "Heat loss" should be reduced as much as possible. "Heat given to the process fluid by preheating" should be maximized, and combustion control should be completely implemented.

For the heating furnace, heating only the process fluid lets the combustion exhaust gas at a high temperature get out. Therefore, this exhaust gas heat is usually used for final preheating of the feed oil or preheating of the combustion air (air preheater) or reheating the steam (super-heater) to attain energy conservation.

This study revealed that most heating furnaces let the combustion exhaust gas of a high temperature get out. For these furnaces, energy conservation can be achieved by installing the air preheater and steam super-heater.

For preheating of the process fluid, the high-temperature process fluid at the downstream of this process is generally used as heat source.

The high-temperature oil in oil refining processes should be cooled down to the specified temperature before the oil is extracted as the intermediate product or final product. If heat to be removed by cooling can be transferred to the feed oil, "required combustion heat" can be reduced. Therefore, a case study of heat utilization optimization with a heat exchanger is required in process design.

If preheating the process fluid is drastically achieved from both aspects of equipment improvement and operation control, "required combustion heat" in the heating furnace can be closer to zero. Actually, a gas oil desulfurization equipment in a plant of Japan has reduced "required combustion heat" in the heating furnace closely to zero (minimum combustion heat in the equipment) by adding a heat exchanger. In this example, if there is an additional small heat source, "required combustion heat" can be zero. In a case of new installation or step III level modification, an equipment without a reheating furnace can be installed.

To implement energy conservation in this refinery, maximum heat recovery by operation control enhancement in step I will be effective for the existing heat exchanger. To enhance heat recovery newly, addition of the heat exchanger in step II or enhancement of measuring equipment will be required.

Generally, an allowance is often added in the design stage considering the property variation of the oil to be processed, in construction of oil refining equipment. When components such as the pump, heat exchanger, CV, etc. are being selected, models having the capacities higher than the designed values are selected. Consequently, excessive capabilities of about 10% are often provided finally.

Thus, the property variation of the oil to be processed should be followed up after startup to optimize the capacity of the equipment, and yet there still remains some room for energy conservation.

(2) Heating furnace

(2-1) Installation status

This refinery has 17 heating furnaces in the North Plant and 22 heating furnaces in the South Plant. Table 2.9 shows the designed values of combustion capacity, raw material processing capacity, etc. of these furnaces. For some heating furnaces, actual data of the combustion volume was available.

Table. 2.9 Heating Furnaces (1/2)

1. North Plant

Tag No.	Unit Name	Design Duty Gcal/h	Operating Duty Gcal/h	Feed Rate kg/h	Outlet Temperature °C	Fuel Kind	Built Year
1.1 ATMOSPHERIC & VACUUM DISTILLATION unit							
2H-101	Atmospheric distillation unit	56.6	51.5	529,900	360	Gas, Oil	1974
2H-102	Kerosene (distillation)	8.9	8.9	172,100	258	Gas	1974
2H-151	Vacuum distillation unit	30.3	28.9	329,700	427	Gas, Oil	1974
2H-181	Lube distillation unit	7.8	7.1	140,200	385	Gas, Oil	1974
1.2 PLATFORMER unit							
2H-201	Unifiner	4.9	2.7	62,130	388	Gas	1974
2H-202	Unifiner stripper	4.3	4.1	135,400	210	Gas	1974
2H-251	Platformer R#1	15.3	11.9	136,500	538	Gas	1974
2H-252	Platformer R#2	12.8	10.7	136,500	538	Gas	1974
2H-253	Platformer R#3	3.2	2.4	136,500	538	Gas	1974
2H-254	Platformer stabilizer reboiler	5.0	5.0	168,500	242	Gas	1974
1.3 VISBREAKER unit							
2H-301	Visbreaking unit	21.2	21.2	126,000	488	Gas	1974
1.4 ISOMAX unit							
2H-401 to 2	Isomax unit	5.7*1	3.9	78,250	381	Gas	1974
2H-403	Isomax unit	9.1	9.1	232,000	323	Gas	1974
2H-404	Isomax unit	9.6	9.6	104,100	371	Gas	1974
2H-405	Isomax unit	0.6	0.6	20,490	318	Gas	1974
1.5 LPG recovery unit							
2H-601	LPG unit	3.8	3.2	99,770	278	Gas	1974
1.6 H_2 gas manufacturing unit							
2H-801	Hydrogen unit	49.9	49.9	7,220	804	Gas	1974

Note *1: Total duty of 2H-401 & 2H-402

Table. 2.9 Heating Furnaces (2/2)

2. South Plant

Tag No.	Unit Name	Design Duty Gcal/h	Operating Duty Gcal/h	Feed Rate kg/h	Outlet Temperature °C	Fuel Kind	Built Year
2.1 ATMOSPHERIC & VACUUM DISTILLATION unit							
H-101	Atmospheric distillation unit	55.5	–	639,300	360	Gas, Oil	1968
H-102	Kerosene (distillation)	8.1	8.1	146,200	258	Gas	1968
H-151	Vacuum distillation unit	24.0	24.0	320,500	391	Gas, Oil	1968
2.2 PLATFORMER unit							
H-201	Unifiner	4.9	3.8	51,620	388	Gas	1968
H-202	Unifiner	3.0	3.0	83,150	199	Gas	1968
H-203	Unifiner	3.0	2.9	97,510	210	Gas	1968
H-251	Platformer R#1	13.9	–	137,500	538	Gas	1968
H-252	Platformer R#2	15.1	11.2	137,500	538	Gas	1968
H-253	Platformer R#3	7.5	–	137,500	538	Gas	1968
H-254	Stabilizer reboiler	6.3	–	185,900	248	Gas	1968
2.3 VISBREAKER unit							
H-301	Visbreaking unit	33.4	30.4	145,000	491	Gas	1968
2.4 ISOMAX unit							
H-430 to 2	Isomax unit	3.0*2	3.0	67,760	413	Gas	1968
H-433	Isomax unit	23.3	–	204,800	391	Gas	1968
2.5 H_2 GAS MANUFACTURING unit							
H-801	Hydrogen unit	28.0	28.0	7,220	804	Gas	1968
2.6 LUBE unit							
H-1101	Propane deasphalting	4.7	–	31,970	260	Gas	1974
H-1102	Propane deasphalting	6.9	–	122,000	260	Gas	1974
H-1201	Furfural unit	1.9	–	25,440	204	Gas	1974
H-1202	Furfural unit	21.0	–	211,200	232	Gas	1974
H-1203	Furfural unit	1.9	–	19,510	204	Gas	1974
H-1301	MEK	6.8	–	59,900	204	Gas	1974
H-1302	MEK	2.5	–	23,050	204	Gas	1974
H1401	Hydrofinish oil	1.5	–	60,200	288	Gas	1974

Note *2: Total duty of H-430, H431 & H432

3. Boiler

Tag No.	Unit Name	Design Duty Gcal/h	Operating Duty Gcal/h	Feed Rate kg/h	Outlet Temperature °C	Fuel Kind	Built Year
B-2101ABCD	Boiler	–	–	145,100	399	Gas, Oil	1968
2B-2101ABC	Boiler	–	–	145,100	399	Gasonline	1974

(2-2) Measurement result and heat balance

a. Outline of measurement

Some heating furnaces in the North Plant and boilers in the South Plant were measured by using the portable measuring equipment carried by the study team. Table 2.10 shows the summary of measurement. For the item measured, oxygen in the exhaust gas, for which no onstream oxygen analyser had been provided, was mainly measured.

Table 2.10 Field Measurement

Equipment	Tag No.	Date	Items
Heating furnace for hydrogen generator	2H-801	16-Jun	Oxygen & temperature of exhaust gas
Atmospheric distillation furnace	2H-101	17-Jun	Oxygen & temperature of exhaust gas
Vacuum distillation furnace	2H-151	18-Jun	Oxygen & temperature of exhaust gas
Heat exchanger	2E-173	19-Jun	Cooling water flowrate
Heat exchanger	2E-258	19-Jun	Cooling water flowrate

All of the heating furnaces measured in this study were of the natural draft type. Since no metering is provided on the utility side such as fuel, fine combustion control is impossible. The measured O_2 content in the combustion exhaust gas is higher than an example of the heating furnace in Japan as shown in Table 2.11 (O_2 content: 2 to 4.2 %, air ratio: 1.1 to 1.25).

The exhaust gas temperature is as high as 320 to 470 °C. The exhaust gas temperature at APH outlet in Japan is 150 to 160 °C, which is close to the acid dew point control value obtained from SO_2.

Table 2.11 O_2 Content in Exhaust Gas and Exhaust Gas Temperature

	O_2 Content in Exhaust Gas	Exhaust Gas Temperature
2H-101	8.4 to 8.6 %	397 °C
2H-151	4.8 %	474 °C
2H-801	10.3 %	320 °C

b. Combustion calculation

For combustion calculation, spreadsheet created on a PC was used. When fuel components, air temperature, oxygen content in exhaust gas, etc. are set, fuel heat generation, combustion gas contents and volume, combustion air volume, air ratio, etc. are calculated and displayed on this spreadsheet. Also, effects of air ratio adjustment and air preheating can be obtained by using oxygen in exhaust gas as a parameter. Table 2.12 shows the combustion calculation result.

Table 2.12 Combustion Calculation

Fuel composition

Component	Volume
CO	1.8 %
CO_2	0.0 %
H_2	6.3 %
CH_4	86.8 %
C_2H_4	2.9 %
C_2H_6	0.0 %
C_3H_8	1.5 %
C_4H_{10}	0.7 %
N_2	0.0 %
O_2	0.0 %
H_2O	0.0 %
Fuel temperature	30

Air condition

Air temperature	30
Ambient temperature	30
Humidity	40 %

Theoretical combustion

Exhaust Gas	Wet	Dry
CO_2	9.6 %	12.0 %
N_2	70.5 %	88.0 %
O_2	0.0 %	0.0 %
H_2O	19.9 %	–
Volume	10.59	8.48

(m^3/m^3-fuel gas)

Air required

Wet	Dry
9.61	9.45

(m^3/m^3-fuel gas)

Heat value of fuel

Hl	8,577
Hh	9,493

(kcal/m^3)

Actual combustion

Exhaust Gas Volume & Composition	Volume	Wet %	Dry %
CO_2	1.02	6.6 %	7.7 %
N_2	11.19	72.7 %	84.8 %
O_2	0.99	6.4 %	7.5 %
H_2O	2.19	14.3 %	
CO	0.00	0.0 %	0.0 %
Total	15.39	100.0 %	100.0 %

(m^3/m^3-fuel gas)

Air consumed

	Wet	Dry
Excess air	4.79	4.71
Air total	14.40	14.16
Air ratio	1.50	1.50

(m^3/m^3-fuel gas)

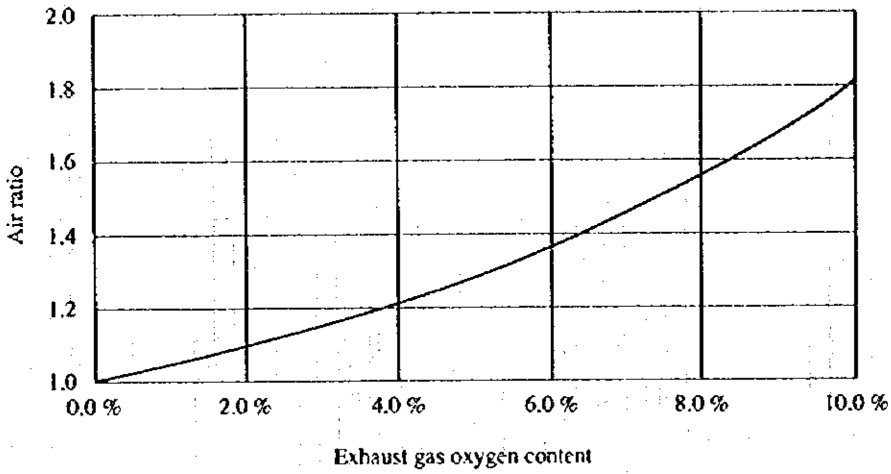
Remarks: Underlined is set value.
Air and gas volume is expressed under normal condition (P = 1.033 kg/cm^2 abs, T = 0 °C)

For air volume and exhaust gas volume and components, the values in both dry and wet states are indicated.

1) Oxygen content in exhaust gas and air ratio

Air ratio is calculated from oxygen content in exhaust gas. Figure 2.8 shows this relationship, which may vary depending on fuel composition, but its effect is small.

Figure 2.8 Air Ratio/Oxygen Content



2) Oxygen content in exhaust gas, exhaust gas volume, and other characteristics

Table 2.13 shows required air volume per fuel unit, exhaust gas volume, CO_2 content, and exhaust gas density in addition to oxygen content in exhaust gas and air ratio

Table 2.13 Characteristics

O_2 (dry) (%)	Air Ratio	Air (wet) (m^3/m^3-fuel gas)	Exhaust Gas (m^3/m^3-fuel gas)	CO_2 (wet) Content in Exhaust (%)	Exhaust Weight (kg/m^3-fuel gas)
0.0	1.000	9.61	10.59	9.6	13.02
2.5	1.121	10.77	11.76	8.6	14.51
5.0	1.281	12.30	13.29	7.7	16.47
7.5	1.499	14.40	15.39	6.6	19.15
10.0	1.816	17.45	18.43	5.5	23.05

c. Crude Oil Heater 2H-101

This equipment is an upright square furnace, on the bottom of which many heavy oil burners are laid out. A radiated heat transfer surface is formed on the inner wall of the furnace with tube walls and horizontal convection heat transfer tubes for oil heating are laid out on the ceiling of the furnace. Combustion gas goes to the stack provided on the top of the furnace via the horizontal tubes and discharged. Air supply to the burners on the bottom of the furnace can be individually adjusted by using the manual damper.

In addition to the general upright furnace structure, this equipment has the steam superheater using horizontal tubes on the ceiling of the combustion chamber.

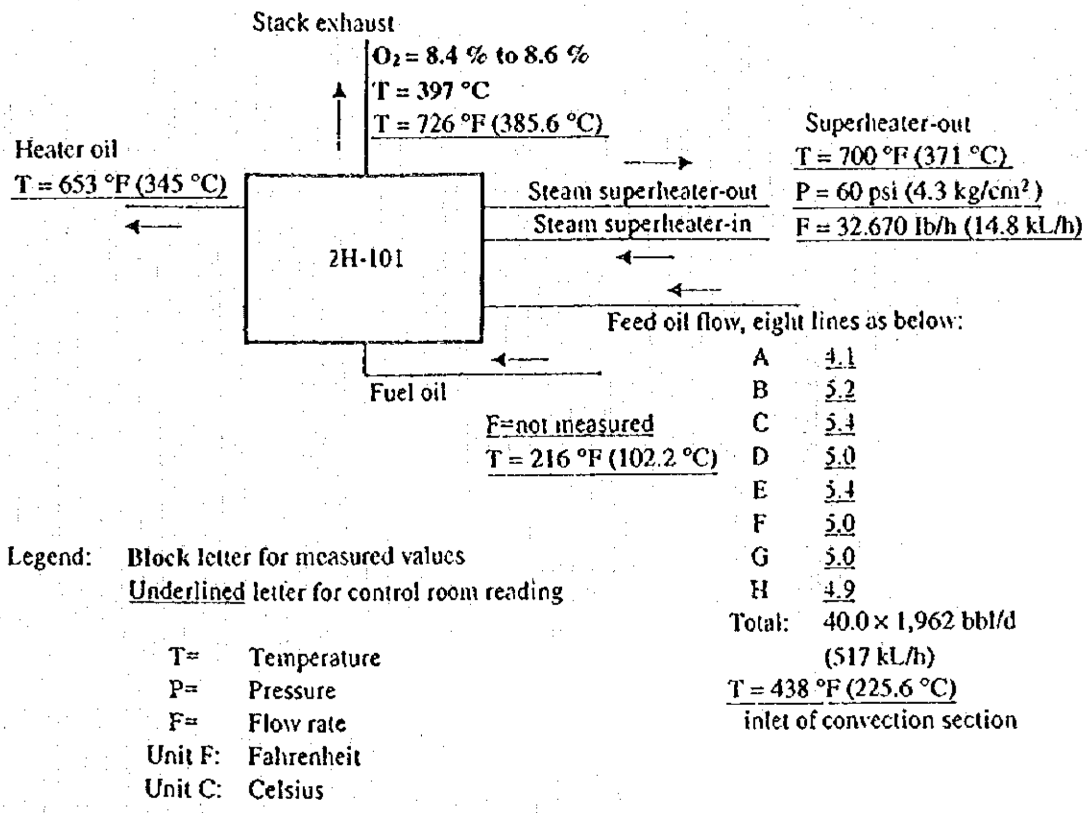
For this equipment, the residual oxygen content and temperature were measured in the exhaust gas flue duct at the inlet of the bottom of the stack. Measurement was connected to the recorder and time-series variation was observed.

1) Measurement result

Figure 2.9 shows oxygen content in exhaust gas and exhaust gas temperature measured in the field and the meter indication in the related control room. In this figure, conversion of the meter indication unit into the metric unit is also shown.

Figure 2.9 Measured Data for Crude Oil Heater 2H-101

<Crude heater 2H-101 for atmospheric distillation, North Plant, on 17 June>



Oxygen content in exhaust gas is not measured in the control room. The exhaust gas temperature indicated by the meter in the control room is approximately identical to the value measured in the field.

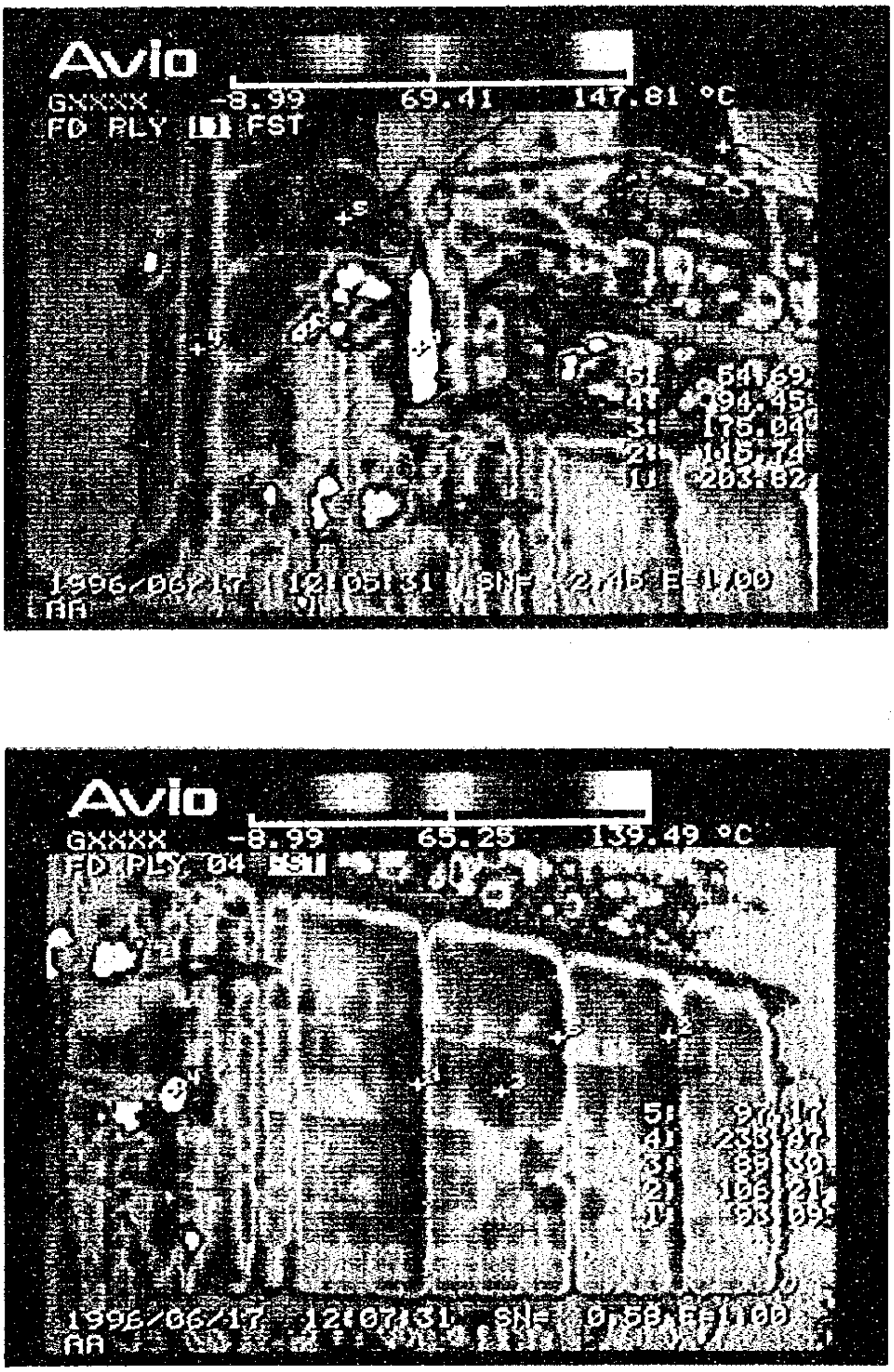
2) Surface temperature and heat emission calculation

As the equipment used to measure the surface temperature of an object in a contactless mode, the "Thermal Video System" is available. With this measuring equipment, infrared emission from the object surface can be caught by a camera and color-displayed on the CRT screen as temperature distribution and the image can be recorded on a floppy disk as a still picture.

This equipment allows viewing the surface temperature of an equipment or steam piping superficially and dynamically and also displaying temperature distribution across any cross section as a graph.

During this field measurement, the external view of the oil heating furnace was photographed with this equipment. Figure 2.10 shows an example of the image.

Figure 2.10 Surface Temperature of Crude Oil Heater 2H-101 by Infrared Visual Display



In this figure, the white color portion is the steam pipe, and its temperature is approximately 200 °C. The surface of the heating furnace body is orange and its temperature is approximately 100 °C.

During field measurement, the temperature of the casing surface on the equipment body was measured at many points by using a contactless radiation thermometer. Heat emission from the surface area on the equipment was calculated by using the measured value. Table 2.14 shows the heat emission calculation result.

Table 2.14 Surface Emission Calculation, 2H-101

Position	Ambient Temperature °C	Surface Temperature °C	Emissivity	Area m^2	Convection Coefficient $kcal/m^2{\cdot}h{\cdot}°C$	Radiation Coefficient $kcal/m^2{\cdot}h{\cdot}°C$	Unit Heat $kcal/m^2{\cdot}h$	Total Heat kcal/h
1st deck	30	103	0.9	576	5.208	6.963	888	511,754
2nd deck	30	148	0.9	144	5.973	8.546	1706	245,672
End plate	30	95	0.9	108	5.030	6.706	763	82,383
Total								839,809
After reinforcement of insulation	30	60	0.9	828	3.957	5.670	289	239,135

The heat emission area was obtained by visual check of the approximate dimensions of the equipment. Since calculation of heat emission is not precise in terms of the measured value, formula, emissivity, etc., it should be understood that the heat emission value is approximate value.

As shown in Table 2.14, when the surface temperature of the furnace body is between 95 °C and 148 °C and the outer air temperature is 30 °C, heat emission from the side wall of the heating furnace and end plate is 839,809 kcal/h. As in Japan, if the surface temperature is kept 60 °C or lower by the improvement of the refractory in the furnace reduces heat emission down to 239,135 kcal/h. This can save, 4,893 Gcal/year (= (839,809 – 239,135) kcal/h × 24 × 365 × 0.93).

3) Heat balance

An attempt is made to calculate the heat balance on this equipment from the measured value and meter indication in the control room. However, since the fuel flow rate has not been measured, the operation heat value shown in the design document is used as the fuel flow rate. Table 2.15 shows the calculation result.

Table 2.15 Heat Balance Calculation on Crude Oil Heater, 2H-101

Heat-in (kcal/h)

	Amount	Unit Heat	Temperature °C	Heat kcal/h	%/Fuel
Feed	517,000 L/h	0.6 kcal/L	225.6	60,675,120	117.9 %
Steam	14,819 kg/h	657.0 kcal/kg		9,735,457	18.9 %[1]
Fuel combustion heat	5,147 L/h	10,002.5 kcal/L		51,483,600	100.0 %[2]
Fuel sensible heat	5,147 L/h	0.438 kcal/kg/°C	102.2	162,918	0.3 %
Total				122,057,095	237.1 %

Heat-out (kcal/h)

	Amount	Unit Heat	Temperature °C	Heat kcal/h	%/Fuel
Product	517,000 L/h	0.6 kcal/L	345.0	97,713,000	189.8 %
Exhaust gas	92,976 m^3_N/h	0.343 kcal/m^3_N/°C	397.0	11,703,916	22.7 %
Steam	14,819 kg/h	767 kcal/kg		11,366,145	22.1 %
Heat release, wall				839,809	1.6 %
Heat release, pipe (30 % of wall, assumed)				251,943	0.5 %
Miscellaneous				182,282	0.4 %
Total				122,057,095	237.1 %
Absorbed heat by product				37,037,880	71.9 %
Absorbed heat by steam				1,630,688	3.2 %

Notes 1: Steam flow quoted from design chart. Steam flow = 32,670 lb/h = 14,819 kg/h
2: Fuel is assumed to be same as Japanese fuel oil of class C.
3: Base temperature: 30 °C

As shown in Table 2.15, approximately 72 % of fuel heat is used to heat the feed oil but 21 % is still released as calories of exhaust gas.

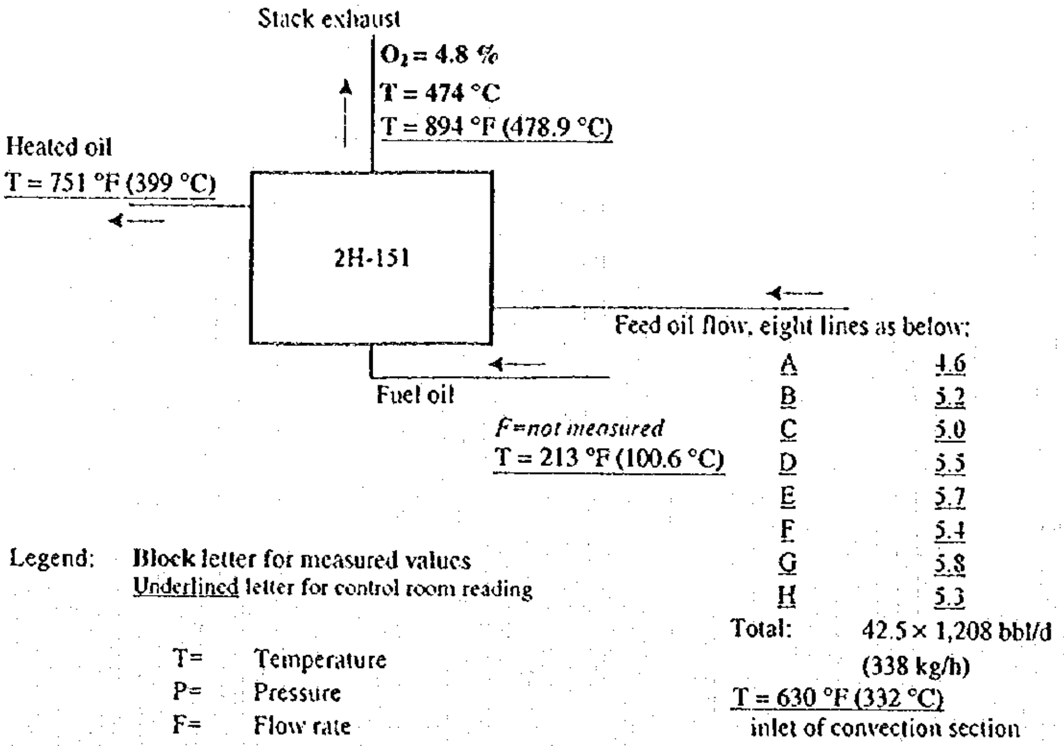
d. Oil heater furnace for the vacuum distillation equipment

This equipment is also an upright type heating furnace. Many heavy oil burners are laid out on the bottom. The inner wall of the furnace consists of tube walls and convection heat transfer tubes are laid out on the ceiling. For field measurement, oxygen content in exhaust gas and exhaust gas temperature were measured at the bottom of the stack.

Figure 2.11 shows the measurement result and meter indication in the related control room.

Figure 2.11 Measured Data for Oil Heater 2H-151

<Oil heater 2H-151 for vacuum distillation, North Plant on 18 June>



Legend: **Block letter for measured values**
Underlined letter for control room reading

T=	Temperature
P=	Pressure
F=	Flow rate
Unit F:	Fahrenheit
Unit C:	Celsius

Oxygen content in exhaust gas is not measured in the control room. For exhaust gas temperature, meter indication in the control room approximately matches the value measured in the field. Oxygen content in exhaust gas was measured to be approximately 4.8 %, which corresponds to 1.26 as the air ratio that is close to 1.25 which is the standard value in Japan. It can be said that air ratio control is excellent. Table 2.16 shows the heat balance calculation result based on the measured value.

Table 2.16 Heat Balance Calculation on Oil Heater, 2H-151

Heat-in (kcal/h)

	Amount	Unit Heat	Temperature °C	Heat kcal/h	%/Fuel
Feed	337,988 L/h	0.6 kcal/L	332.0	61,243,426	211.7 %
					0.0 %
Fuel combustion	2,892 L/h	10,002.5 kcal/L		28,929,600	100.0 %[1,2]
Fuel sensible heat	2,892 L/h	0.438 kcal/kg/°C	100.6	89,518	0.3 %
Total				90,262,543	312.0 %

Heat-out (kcal/h)

	Amount	Unit Heat	Temperature °C	Heat kcal/h	%/Fuel
Product	337,988 L/h	0.6 kcal/L	399.0	74,830,543	258.7 %
Exhaust gas	41,113 m^3_N/h	0.348 kcal/m^3_N/°C	474.0	6,352,439	22.0 %
Emission, wall				471,944	1.6 %[3]
Emission, pipes (30 % of wall, assumed)				141,583	0.5 %
Miscellaneous				8,466,034	29.3 %
Total				90,262,543	312.0%
Absorbed heat by product				13,587,118	47.0 %

Notes 1: Fuel flow is quoted from design chart because no meter is equipped in control room.
2: Fuel is assumed to be same as Japanese fuel oil of class C.
3: Emission was assumed to be proportional to heating capacity to crude heater 2H-101.
4: Base temperature: 30 °C

For this calculation, the item (fuel flow rate) for which any measured value was unavailable was quoted from the design document. Heat emission was obtained as an approximate value proportional to the heat input from heat emission from the oil heater 2H-101. According to the calculation result, 47 % of fuel heat is effectively used as shown in Table 2.16.

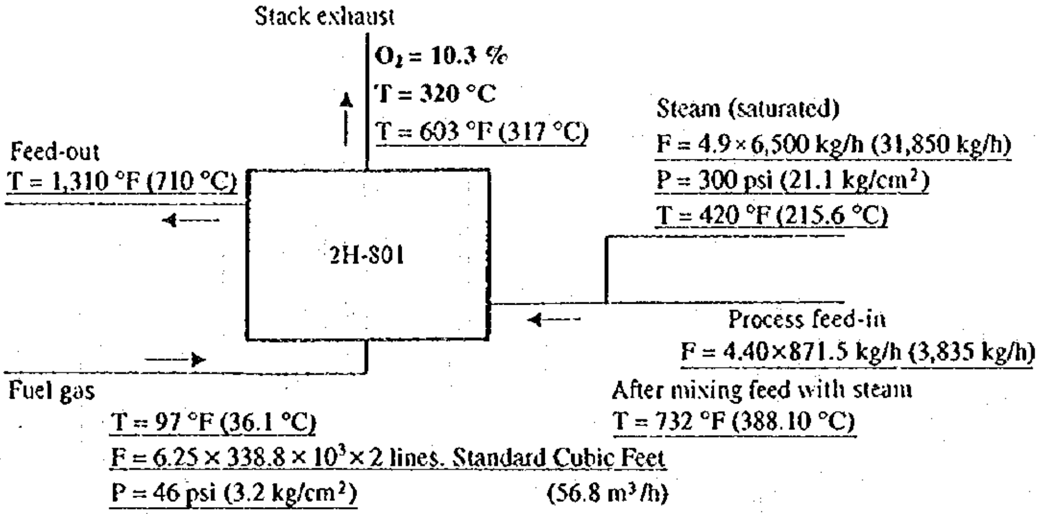
e. Heating furnace for hydrogen generator 2H-801

This equipment is an upright furnace and has many gas burners laid out on the bottom. For field measurement, oxygen content in exhaust gas and exhaust gas temperature were measured in the duct at the bottom of the stack.

Figure 2.12 shows the measurement result and meter indication in the related control room.

Figure 2.12 Measured Data for Hydrogen Generator 2H-801

<H-801, Hydrogen generator, on 16 June>



Legend: **Block letter for measured values**
Underlined letter for control room reading

T= Temperature
P= Pressure
F= Flow rate

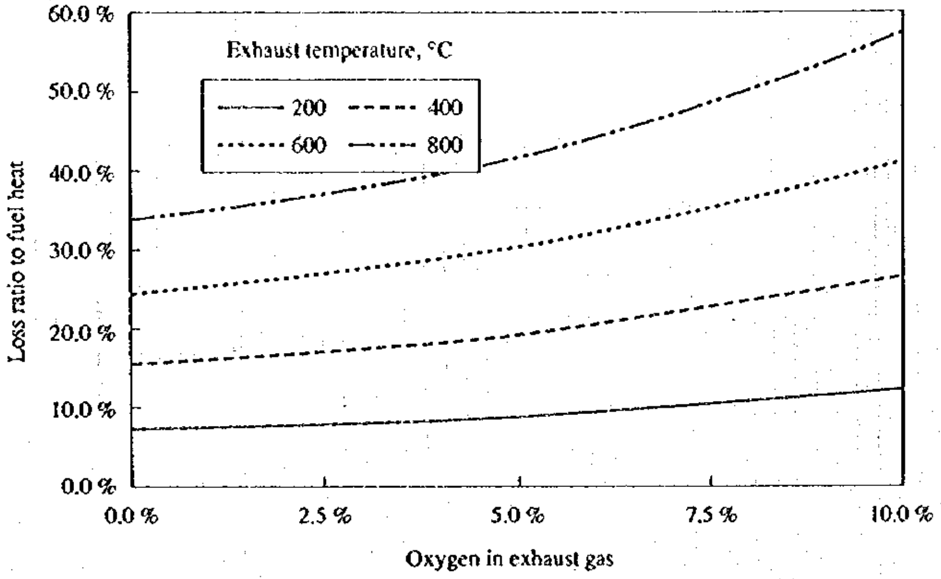
As shown in this figure, oxygen content in exhaust gas is 10.3 %, which corresponds to 1.86 as the air ratio. This value is much higher than 1.25 that is the standard air ratio in Japan and it can be said that there is a large exhaust gas loss. If oxygen in exhaust gas is reduced from 10.3 % as the measured value to 4 % which corresponds to air ratio 1.25, exhaust gas loss reduces from 20.2 % to 14.2 % according to combustion calculation. Fuel consumption by adjustment of air ratio is reduced by 7.7 % according to calculation.

(2-3) Reduction of exhaust gas loss

a. Exhaust gas loss

A large air ratio means that air more than required is heated and discharged. Therefore, as the air ratio is larger, heat loss by exhaust gas increases. Figure 2.13 shows the exhaust gas loss against fuel input heat by using exhaust gas temperature as a parameter in the relationship with oxygen content in exhaust gas.

Figure 2.13 Exhaust Gas Loss – North Plant Gas Firing



As shown in Figure 2.13, reducing the oxygen content in exhaust gas decreases heat loss by exhaust gas. The degree of reduction is larger when the exhaust gas temperature is higher. Table 2.17 shows an example of calculation for fuel reduction by decreasing air ratio.

As shown in this table, fuel reduction on the vertical axis can be known by the oxygen content in exhaust gas before improvement and the one after improvement.

Table 2.17 Fuel Economy by Air-ratio Adjustment (Gas Firing)

Exhaust gas temperature = 500

Before adjustment	O_2 dry after adjustment			
O_2 dry	0.0 %	2.5 %	5.0 %	7.5 %
0.0 %	0.0 %	–	–	–
2.5 %	2.6 %	0.0 %	–	–
5.0 %	6.0 %	3.5 %	0.0 %	–
7.5 %	10.6 %	8.2 %	4.9 %	0.0 %
10.0 %	17.3 %	15.1 %	12.1 %	7.5 %

Exhaust gas temperature = 600

Before adjustment	O_2 dry after adjustment			
O_2 dry	0.0 %	2.5 %	5.0 %	7.5 %
0.0 %	0.0 %	–	–	–
2.5 %	3.3 %	0.0 %	–	–
5.0 %	7.7 %	4.5 %	0.0 %	–
7.5 %	13.7 %	10.8 %	6.5 %	0.0 %
10.0 %	22.5 %	19.8 %	16.0 %	10.1 %

Exhaust gas temperature = 700

Before adjustment	O_2 dry after adjustment			
O_2 dry	0.0 %	2.5 %	5.0 %	7.5 %
0.0 %	0.0 %	–	–	–
2.5 %	4.2 %	0.0 %	–	–
5.0 %	9.8 %	5.8 %	0.0 %	–
7.5 %	17.4 %	13.7 %	8.4 %	0.0 %
10.0 %	28.4 %	25.3 %	20.7 %	13.4 %

Exhaust gas temperature = 800

Before adjustment	O_2 dry after adjustment			
O_2 dry	0.0 %	2.5 %	5.0 %	7.5 %
0.0 %	0.0 %	–	–	–
2.5 %	5.3 %	0.0 %	–	–
5.0 %	12.2 %	7.3 %	0.0 %	–
7.5 %	21.6 %	17.3 %	10.8 %	0.0 %
10.0 %	35.4 %	31.8 %	26.4 %	17.6 %

This calculation assumes that the exhaust gas temperature does not change when oxygen content in exhaust gas changes and that heat other than exhaust gas loss is the required one and not changed by oxygen content in exhaust gas.

The power required for the forced draft fan and the induced draft fan also decreases.

b. Air ratio control

1) Management enhancement

If O_2 content in exhaust gas is completely controlled by maintaining operation manuals and increasing frequency of adjusting the combustion air volume although the equipment is not improved, excessive air in the current status can be reduced. However, since oxygen content in exhaust gas cannot be checked without the oxygen meter, combustion control requires 5 % to 7 % of oxygen content and 1.3 to 1.5 of air ratio for safety.

2) Combustion control using the portable oxygen meter

For promotion of energy conservation, the minimum measuring equipment is required. For combustion management in step II, one or two simple, portable oxygen meters (¥ 300,000 for two sets) should be prepared by a minimum investment. The management method that sequentially measures oxygen content in each heating furnace in several days or when the operating condition is changed is effective.

In this step, operation in a level of 4 to 5 % of oxygen content and 1.2 to 1.3 of air ratio is allowed. As excessive oxygen content is lower, measurement accuracy becomes more important. Therefore, when oxygen content is actually being reduced, air incoming from the furnace wall, etc. must be completely prevented. To make operators participate in energy conservation, it is important to perform these activities by themselves.

3) Automatic combustion control by the installation of oxygen meter

Automatic combustion control combined with FDF (forced draft fan) and IDF (induced draft fan) using a permanently installed exhaust gas oxygen meter can be considered. The basic of automatic combustion control is feedback control with the oxygen meter but feed-forward control including variation of the volume of feed oil can also be considered. As oxygen content in exhaust gas reduces, partial incomplete combustion may result upon operation variation. However, combustion control may be enhanced by applying the CO and CO_2 meters for preventing incomplete combustion.

For these automatic controls, combustion in a level of 2 to 3 % of oxygen content and 1.1 to 1.2 of air ratio or in a lower level is allowed.

As combustion control is intensified, securing the oxygen content measuring accuracy becomes more important. Therefore, air incoming from the furnace wall, etc. must be prevented in this step also. To enhance combustion control, operators must apply inflammable heat insulation material into the gap or opening between the heating tube and furnace wall by themselves to prevent air incoming so that they will be conscious of participation in energy conservation activities.

c. Air preheating

The fuel volume can be reduced if combustion air is preheated by using the heat of exhaust gas. Fuel reduction by air preheating can be calculated by using heat balance calculation. Supposing that combustion air is heated to 300 °C by combustion exhaust gas when oxygen content in exhaust gas is 7.5 % and exhaust gas temperature is 400 °C. Then, fuel reduction is 15.6 % against the fuel volume required before preheating. The fuel reduction rate is higher as the preheated air temperature increases, and oxygen content in exhaust gas is larger. Table 2.18 shows the calculation result under various conditions.

Table 2.18 Fuel Economy by Air Preheating

Exhaust gas temperature before preheating = 400

O_2 dry	Preheated Air Temperature 300	400	500
0.0 %	10.2 %	–	–
2.5 %	11.5 %	–	–
5.0 %	13.2 %	–	–
7.5 %	15.6 %	–	–
10.0 %	19.1 %	–	–

Exhaust gas temperature before preheating = 500

O_2 dry	Preheated Air Temperature 300	400	500
0.0 %	10.7 %	14.2 %	–
2.5 %	12.1 %	16.0 %	–
5.0 %	14.0 %	18.4 %	–
7.5 %	16.7 %	21.7 %	–
10.0 %	20.8 %	26.6 %	–

Exhaust gas temperature before preheating = 600

O_2 dry	Preheated Air Temperature 300	400	500
0.0 %	11.3 %	14.9 %	18.3 %
2.5 %	12.8 %	16.9 %	20.7 %
5.0 %	15.0 %	19.6 %	23.8 %
7.5 %	18.1 %	23.3 %	28.1 %
10.0 %	22.9 %	29.1 %	34.5 %

Exhaust gas temperature before preheating = 700

O_2 dry	Preheated Air Temperature 300	400	500
0.0 %	11.9 %	15.7 %	19.3 %
2.5 %	13.7 %	17.9 %	21.9 %
5.0 %	16.1 %	21.0 %	25.4 %
7.5 %	19.7 %	25.3 %	30.3 %
10.0 %	25.5 %	32.2 %	37.8 %

d. Effect of improvement of air ratio and air preheating

For combustion in each furnace, on assumptions of air ratio improvement and air preheating by exhaust gas, a trial calculation of the allowance for energy conservation for the entire refinery is made.

The following assumptions are made for trial calculation:

1) The exhaust gas temperature is approximately 200 °C higher than the temperature at the furnace outlet for the raw material to be heated. Oxygen content in exhaust gas before improvement is 8 %.
These assumptions are made for convenience in calculation on the computer and may be different from real data. However, there is no problem when accuracy of calculation is considered.

2) Air ratio after improvement is 1.25 as the standard value in Japan. This value corresponds to 4 % oxygen in exhaust gas.

3) For air preheating, the temperature of air to be preheated is 200 °C lower than the exhaust gas temperature. In this case, the temperature of exhaust gas at the outlet of the air preheater is approximately 200 °C.

4) Heat emission from the air preheater and leak from the air side to the gas side are not considered.

5) Assuming that air preheating is performed after air ratio improvement, calculation is made based on 4 % of oxygen in exhaust gas.

Table 2.19 shows the calculation result based on these assumptions. This table lists the calculation results before improvement, those after air ratio improvement, and those after air preheating, each of which is provided with the fuel consumption reduction rate.

Table 2.19 Estimated Fuel Saving by Air Ratio Adjustment & Air Preheating

North Plant Tag No.	Facility Name	Operating Gcal/h	Fuel	<Before improvement> Exhaust Temperature	Exhaust Gas Loss	Exhaust Loss Total Fuel	<Air Ratio Adjusting> Advantage (8 % to 4 %)	Save to Total Fuel	<Air Pre-heating> Advantage (O_2 = 4 %)	Save to Total Fuel
Atmospheric/Vacuum distillation unit										
2H-101	Atmospheric distillation unit	51.48	Gas, Oil	397	22.7 %	5.07 %	6.4 %	1.42 %	7.8 %	1.73 %
2H-102	Kerosene (distillation)	8.91	Gas	500	29.3 %	1.13 %	7.6 %	0.29 %	13.2 %	0.51 %
2H-151	Vacuum distillation unit	28.93	Gas, Oil	474	22.0 %	2.76 %	1.1 %	0.14 %	11.9 %	1.49 %
2H-181	Lube distillation unit	7.06	Gas, Oil	500	27.7 %	0.85 %	7.6 %	0.23 %	13.2 %	0.40 %
Unifier										
2H-201	Unifiner	2.72	Gas	600	36.0 %	0.42 %	10.0 %	0.12 %	18.4 %	0.22 %
2H-202	Unifiner stripper	4.10	Gas	600	36.0 %	0.64 %	10.0 %	0.18 %	18.4 %	0.33 %
2H-251	Platform R#1	11.89	Gas	800	49.7 %	2.57 %	16.3 %	0.84 %	29.5 %	1.52 %
2H-252	Platform R#2	10.68	Gas	800	49.7 %	2.30 %	16.3 %	0.75 %	29.5 %	1.37 %
2H-253	Platform R#3	2.42	Gas	800	49.7 %	0.52 %	16.3 %	0.17 %	29.5 %	0.31 %
2H-254	Platform stabilizer reboiler	5.00	Gas	400	22.8 %	0.50 %	5.5 %	0.12 %	8.1 %	0.18 %
Visbreaker unit										
2H-301	Visbreaking unit	21.19	Gas	700	42.8 %	3.93 %	12.8 %	1.18 %	23.8 %	2.19 %
RCD isomax unit										
2H-401	Isomax unit	3.87	Gas	600	36.0 %	0.60 %	10.0 %	0.17 %	18.4 %	0.31 %
2H-402										
2H-403	Isomax unit	9.08	Gas	600	36.0 %	1.42 %	10.0 %	0.39 %	18.4 %	0.72 %
2H-404	Isomax unit	9.61	Gas	600	36.0 %	1.50 %	10.0 %	0.42 %	18.4 %	0.77 %
2H-405	Isomax unit	0.56	Gas	600	36.0 %	0.09 %	10.0 %	0.02 %	18.4 %	0.05 %
LPG refining unit										
2H-601	LPG unit	3.23	Gas	500	29.3 %	0.41 %	7.6 %	0.11 %	13.2 %	0.18 %
Hydrogen generator										
2H-801	Hydrogen unit	49.84	Gas	400	20.2 %	4.37 %	7.7 %	1.67 %	4.1 %	0.88 %
	Total	230.59				29.08 %		8.23 %		14.28 %

Notes: Exhaust temperature is categorized into round number considering feed-out temperature except for measured furnaces.
Existing oxygen content is set to 8 % for all furnaces except measured furnaces.
Preheated air temperature is selected so that the temperature difference of air and gas at the hot end of air-heater will be 200 °C.
The above temperature makes airheater-out gas temperature around 200 °C.
Categorizing of gas/air and oxygen content is necessary to make the calculation simple on PC worksheet.

According to this calculation, exhaust gas loss before improvement is 30 % for the entire refinery. As a result of air ratio improvement, fuel consumption is reduced by 8.23 % (154,546 Gcal/year (= $230.59 \times 10^6 \times 0.0838 \times 8,760 \times 0.93$)). And remaining fuel consumption is further reduced by 14.28 % (246,179 kcal/year (= $(230.59 \times 106 - 157.424) \times 0.1428 \times 24 \times 365 \times 0.93$)) by the air preheating.

As a result of calculation, the energy conservation potential relating to combustion in the heating furnace is indicated. Air ratio improvement can be implemented with a relatively small equipment cost. On the other hand, the following problems may have to be examined for installation of the air preheater:

1) It is necessary to install a large structure on the top of the furnace (i.e. bottom of the stack).

   Can a large-diameter duct which guides the high-temperature preheated air from the air preheater on the top of the furnace to the burners on the bottom of the furnace be installed?

2) On the bottom of the furnace, the preheated air should be distributed equally to many burners.

3) Existing burners may not accommodate utilization of the preheated air, so they should be replaced.

The same situation applies to the boilers. The measured exhaust gas temperature of the boilers in the North Plant is said to be 390 to 450 °C.

(2-4) Heating pipe path balance control

For a heating furnace in which multiple heating pipe paths are provided, uniform heating of the paths is essential for long-term stable operation suppressing coking (i.e. energy conserving operation). In this study, it was found that the ratio between the maximum and minimum on each path balance in heating furnaces 2H-101 and 2H-151 was approximately 1.3, which is large, and the opening degrees of burners' air dampers were uneven.

2H-101: FRC-103 A-H 4.1 to 5.3 (×1,962 BPSD)
2H-151: FRC-165 A-H 4.5 to 5.7 (×1,208 BPSD)

According to operators, this situation was caused by adjustment depending on the differences in the in-furnace position of heating pipes and burners and heating pipe form for each path. However, since the feed oil is mixed and fed, each path CIT (Coil Inlet Temperature at the inlet of the heating furnace) is always same. Therefore, the paths should be operated under the same heating conditions.

If such unbalanced operation is continued, coking may lead to difficulty in long-term stable operation for paths with a low flow rate (i.e. low in-pipe flow speed) of crude oil.

(2-5) H_2 manufacturing plant heating pipe TMT management

Operation of the heating furnace in the H_2 manufacturing plant is adjusted with a low frequency (once a month). During the high-load operation or when the combustion status is poor (i.e. fuel and air ratio unbalanced), the value of the heating pipe TMT (Tube Metal Temperature) measured by using the pyrometer rises to 1,090 to 1,200 °C. The designed temperature of the heating pipe is 810 to 980 °C, the measured TMT exceeding the designed value implies hazardous operation. However, measurement with the pyrometer tends to result in errors, so accuracy should be improved by taking the following actions:

For the heating furnace not provided with thermocouples at critical points of the heating pipe for TMT measurement, TMT is indirectly measured by using a pyrometer. Use of the pyrometer alone results in a large error and the measured values are for reference only.

Recent TMT measuring methods employed in Japan are:

1) Pad/Tip type: A pad or tip metal piece is attached to the surface of the heating pipe. A sheathed thermocouple is contacted and attached to the pad or tip.

2) Protection cover type: The pad and thermocouple or the single-piece sheathed thermocouple alone are covered with a protection cover (or protection tube) and attached to the heating pipe.

3) Knife-edge type: The tip of a sheathed thermocouple is machined to a knife-edge type and then, the thermocouple is attached to the heating pipe.

Among these, the pad type or tip type measuring method tends to indicate a higher temperature due to the effect of the contact error, etc., while the protection cover type tends to indicate a lower temperature due to the heat transfer problem. Since the knife-edge type has a simple shape and satisfactory accuracy, many users are using it in Japan. In any case, the thermocouple itself experiences aging, so long-term transition follow-up with a combination with the pyrometer is effective.

If normal correlation between the TMT and pyrometer indication is seized, judgment can be made precisely when TMT changes abruptly upon occurrence of a hot spot on the heating pipe.

In this refinery, necessity and applications of TMT management will be important along with promotion of energy conservation (e.g. heating furnace TMT control for the H_2 manufacturing equipment, heating furnace hot spot prevention and corrective action, and heating furnace TMT control during VGO deep-cut operation of the vacuum distillation equipment).

For a while, each burner's combustion air should be adjusted immediately after the operation load changes to avoid the TMT rise caused by unbalanced combustion by maintaining proper operation manuals.

(2-6) Burner coking prevention

During the hearing survey, it was reported that the burner tip is blocked due to coke adhered on some burners in the heating furnace. Actions to be taken, based on the hearing, are as follows and they are indispensable to burners' stable combustion:

1) Prevention of condensate mixing into atomizing steam
(Installation or reinforcement of steam traps at critical points of steam piping)

2) Complete control of ΔP (pressure difference) for fuel oil and steam
(Preparation of operation manuals and enhancement of operation control)

3) Uniformity of fuel properties
(Oil properties (MW and Vis.) may be changed by kinds or blending ratio of slop oil.)

Review of the entire fuel system by categorizing the fuel sources and changing the combination (i.e. slop oil, etc. that may result in troubles are treated individually).

(3) Maximum heat exchange operation

a. U_0 value management

In the refinery, many heat exchange components such as the crude oil preheating heat exchanger, heating pipes of the boiler and heating furnace, and waste heat recovery equipment are provided. To implement the maximum heat exchange (energy conserving) operation by fully utilizing the capabilities of these equipment, each overall heat transfer coefficient (U_0; kcal/m²h °C) should always be calculated and followed up. If U_0 tends to decrease, it is necessary to maintain efficiency by adjusting the operation immediately or cleaning the heat transfer surface when required.

b. Path balance control

In this refinery, there are partially two lines of heat exchangers for preheating the crude oil to the atmospheric distillation unit. During normal operation, path balance should be controlled by adjusting the crude oil flow rate to each line when the operating condition is changed so that the heat exchange volume in each line will be the maximum.

This study revealed that the designed value for the crude oil dividing ratio for 2E-156, 2E-182, and 2E-167 was 40 % but the actual value was 70 %, which is large. The background on this problem was not checked. Daily efforts for fine operation adjustment for maximum heat exchange are required.

c. Enhancement of heat recovery from the cooler

According to the hearing from the plant people, there was no particular standard for selecting the use of the CTW cooling water circulated type cooler or air fin cooler in the design stage. However, in general cooler design, because of the approach temperature difference (ΔT) between the process fluid and cooling medium, the air fin type should be used for the high-temperature fluid and the cooling water circulated type should be used for the low-temperature fluid. In this way, the cooling water circulating equipment can be made smaller.

Therefore, from the viewpoint of energy conservation, the air fin cooler installed at the relatively high temperature side, rather than the cooling water circulated type, has possibility of heat recovery improvement.

From the process flow sheet, possibility of enhancement in heat recovery to crude oil was examined for the coolers of kerosene and light diesel in the atmospheric distillation equipment and the coolers of isomax feed and vacuum bottom in the vacuum distillation equipment. Table 2.20 shows the design temperature and actual operation temperature of these coolers.

Table 2.20 Design Base & Actual Data of Coolers

Tag No.		2E-106 A&B	2E-111 A&B	2E-107 A&B	2E-109	2E-112 A&B
Service		Kerosene cooler	Kerosene cooler	Light diesel cooler	Light diesel cooler	Light diesel cooler
Surface area	m^2	122	45.3	116		42.6
Cooling duty	Gcal/h	7.11	1.93	7.51	2.68	0.74
Process Fluid		Kerosene	Kerosene	Light diesel	Light diesel	Light diesel
Flow rate	kL/h	407	407	141	141	141
In/out temperature	°C	247/92.2	92.2/40.6	247/116	116/60.0	60.0/43.3
Actual in/out temperature	°C		- /62.2		- /42.8	
Cooling fluid		Crude	CTW	Crude	Air	CTW
Flowrate	BPSD	100,500		100,500		
In/out temperature	°C	60.0/85.6	29.4/46.1	85.5/108		29.4/46.1
Actural in/out temperature	°C	66.7/91.1		91.1/108		

Tag No.		2E156 A&B	2E-162 A&B	2E-159 A-D	2E-173
Service		Isomax feed to storage	Isomax feed to storage	Vacuum bottom cooler	Vacuum bottom cooler
Surface area	m^2	115		293	317
Cooling duty	Gcal/h	2.55 to 6.63	1.51 to 2.38	8.95 to 9.23	5.04
Process Fluid		Isomax feed	Isomax feed	Vacuum bottom	Vacuum bottom
Flow rate	kL/h	99,626	99,626		
In/out temperature	°C	249/149	149/93.3	277/193	193/128
Actual in/out temperature	°C		- /76.7		- /125
Cooling fluid		Crude	Air	Crude	CTW
Flowrate	BPSD			91,254	
In/out temperature	°C	108/144		188/200	29.4/46.1
Actural in/out temperature	°C				

For each cooler, the actual operation temperature data was hardly available and details cannot be examined. According to the process fluid temperature on a design base, totally 2 Gkcal/h heat recovery is possible for 2E-109, 2E-162, and 2E-173 as shown in Table 2.21.

To implement the heat recovery study actually, optimization including rearrangement of existing heat exchangers will be required.

Table 2.21 Heat Recovery from Coolers

Tag No.	Cooling Fluid	Duty (Gcal/h)	Process Fluid Temperature Inlet/outlet (°C)	Modification Cost (k¥)
2E-109	Air	0.7	116/60	17,000
2E-162	Air	0.4 to 0.6	149/93	15,000
2E-173	CTW	1.3	193/128	30,000

(4) Process catalyst control

To implement energy conservation operation in oil refining, it is extremely important to control the process catalyst activity and its residual life. If the catalyst activity is fully utilized to maximize the product throughput with the limited equipment capability within the limited operation period, energy conserving operation with high production efficiency can be achieved.

a. Reformer catalyst activity control

In refineries in Japan, reformer catalyst activity control is one of the most important energy conservation control items in terms of time productivity improvement. Therefore, during normal operation, the engineering group implements the case study relating to the optimum RON (research octane number) level of naphtha always taking the product throughput into account and the proper catalyst regeneration period.

Generally, when activity of the reformer catalyst reduces, the H_2 content in the off gas gradually reduces by time and content of the light HC gases (C_1, C_2, C_3, etc.) rises.

If transitions of the H_2 content in the off gas and content of light HC gases are followed up, then CPI (catalyst performance index), etc. are followed up by using such physical property data, the time required to regenerate the catalyst can be forecast. At the same time, production efficiency can be increased by adjusting operation severity (e.g. processing rate, hydrogen circulation rate, reaction temperature, etc.) such as forecasting the catalyst's residual life (residual operable time) and increasing RON of product naphtha so that there will be no residual life.

For CPI, the process licenser normally owns the formula and, for this refinery, it may be impossible to obtain the formula.

To follow up the reformer catalyst activity, this refinery implements off gas analysis every two weeks and sends the analysis result to UOP, the licenser, for judgment of the catalyst regeneration time. However, without the licenser's CPI, catalyst activity can be controlled by following up the off gas composition transition on a time-series basis.

Since this refinery has a large amount of real operation data and analysis data that have been collected since its startup, the residual life study matching the equipment, that is more precise than the licenser's formula, will be allowed if an original correlative formula is created by using such data.

Regular maintenance in these plants is performed every three years. So far, this refinery has regenerated the catalyst every 2.5 to 3 years. It may be desirable in terms of reducing the catalyst regeneration cost. However, for maximum production and high value of high RON blend stock for manufacturing gasoline, operation with increased severity (e.g. so that catalyst will be regenerated after about 1.5 years) may result in total merits.

In refineries in Japan, regular maintenance is conducted every two years. For maximum production of the high RON blend stock, high severity operation is often implemented to regenerate the reformer catalyst every year. For this refinery, the optimization case study regarding this matter may be necessary.

b. Desulfurization catalyst activity control

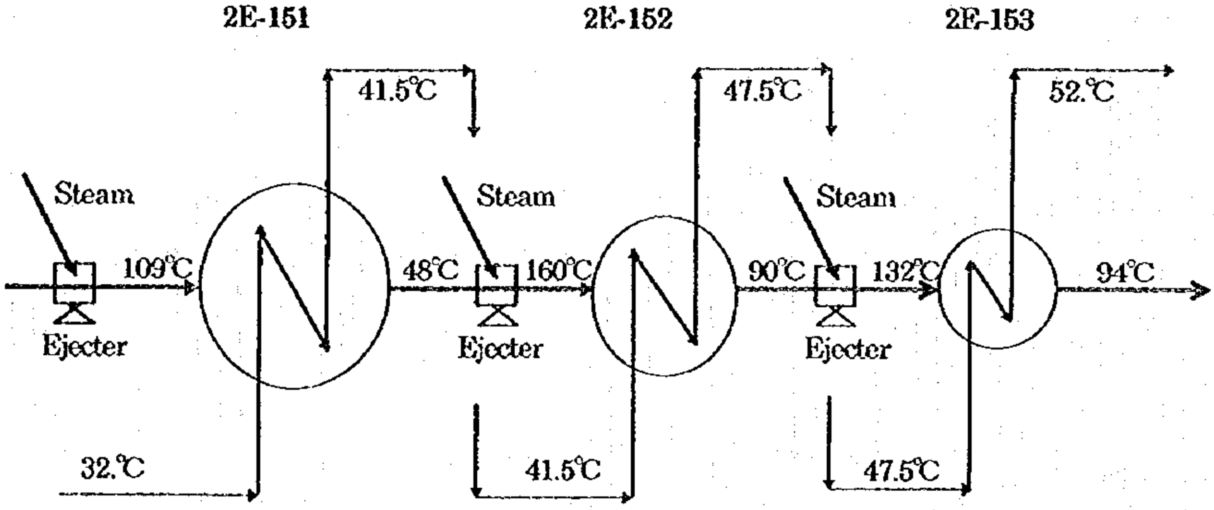
In the hydrodesulfurization equipment such as the isomax unit, energy intensity largely increases when the desulfurization level is increased. Therefore, energy conservation is allowed by controlling the desulfurization level within the optimum range to the extent the product specifications are fulfilled from the aspects of the running cost and catalyst cost (for regeneration, replacement, etc.) The product property analysis group, operation group, and control group should collaborate with each other so that the product quality will not be excessively higher than the specifications.

(5) Improvement of the vacuum distillation overhead cooler

Measurement of the temperature around the vacuum distillation overhead cooler in this study revealed that the cooling capability of overhead gas coolers (2E-151, 2E-152, 2E-153) was poor. Although the gas temperature at the 2E-151 outlet on the first stage is 48 °C, it abruptly rises to 90 °C at the 2E-152 outlet on the second stage then it is 94 °C at the 2E-153 outlet on the third stage, which is much higher than the designed value (38 °C). The cooling water supply temperature is 32 °C, which is slightly higher than the designed value (26.7 °C), but it does not increase the 2E-153 outlet temperature to 94 °C. Figure 2.14 shows the measured temperature balance around these coolers.

Figure 2.14 Temperature Profile of Vacuum Overhead Cooler

Actual Parallel Flow Case



For the vacuum distillation overhead system, a multi-stage (three-stage for this equipment) steam ejector is generally used to maintain vacuum. At the same time, the gas volume is reduced and vacuum distillation operation is accomplished when the oil vapor in the overhead gas is cooled by the cooler and condensed and removed. For the current temperature balance of this equipment, the oil vapor condensation effect by cooling is insufficient for the second cooler onward.

Even under such circumstances, operation mainly with the vacuum effect by the steam ejector is anyway possible. However, efficient vacuum distillation that maximizes the VGO throughput cannot be achieved because the condensation effect is low. The actual detailed cause for insufficient cooling is unknown. Currently, excessive steam may be supplied to the ejector to compensate insufficient cooling and maintain the vacuum effect on the overhead line. As a result, heat of the supplied steam may not be removed by the second cooler onward.

According to the hearing from the plant people, the vacuum distillation tower cannot exhibit the specified capability due to leak from the upper tray and the vacuum top internal packing will be replaced to improve the capability (i.e. increase VGO). However, the cause for the specified capability not being exhibited may possibly be the insufficient cooling capability of the overhead system. For this equipment, the insufficient cooling capability of the overhead system, rather than replacement of the packing, should have the top priority.

In this study, sufficient data could not be collected. It is a matter of immediate concern to collect the detailed design data and actual operation data, identify problems, and take proper actions. Possible actions are:

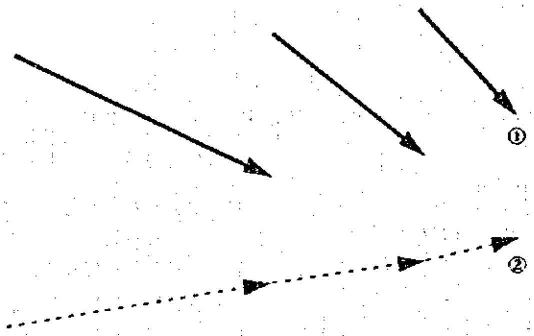
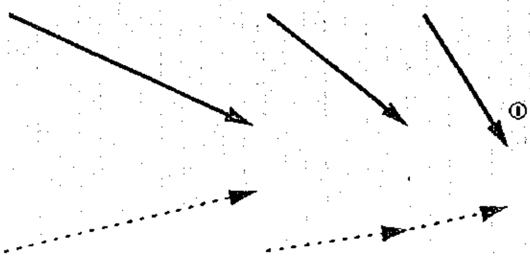
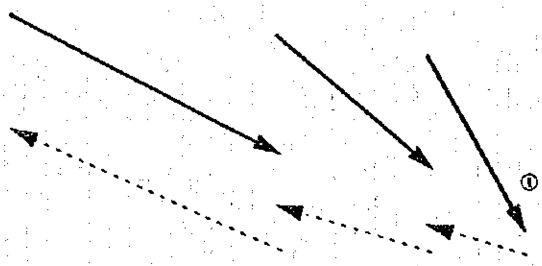
a. Check the basic operating conditions such as the normal operation load, overhead vacuum, steam injection rate to the ejector, etc. and adjust any item deviating from the design condition to the proper level.

b. Check the heat transfer coefficient of three overhead coolers and clean the heat transfer surface as required.

If problems still remain, take the following action:

c. Currently, the overhead gas and cooling water flow from 2E-151 to 2E-153 as one-line concurrent flow. As shown in Table 2.22, the flow should be changed into parallel lines of flow to enhance the cooler's cooling capability.

Table 2.22 Cooling Effect of CTW Flow Change

Case	Approximate Flow	Characteristics
Current flow		• At present, the flow of the tower top pressure gas and the cooling water for 3 coolers (3-stage cooling) is one line concurrent. Therefore it is difficult to reduce the difference between the gas outlet temperature ① and the cooling water outlet temperature ②.
Improvement proposal A flow		• Gas outlet temperature ① can be reduced by supplying the cooling water in 2 lines of parallel flow. • The cooling effect in this case can be improved, but nearly two times more cooling water will be required.
Improvement proposal B flow		• The gas outlet temperature ① can be reduced to the minimum possible point by supplying the cooling water in 3 lines of parallel flow. (In this case there will be no distinction between the concurrent flow and the counter-current flow) • In this case the cooling effect is the largest, but nearly 3 times more cooling water will be required.)

In addition, in Japan, as the actions for efficiency improvement (i.e. deep-cut) of the vacuum distillation tower, reduction of the cooling water temperature (improvement of the cooling capability) by introducing a chiller, installation of the pre-cooler, and installation of a vacuum pump at the last stage of the steam ejector (improvement of vacuum) are employed.

As a step following recovery of the original capability of the vacuum distillation equipment by the cooling water piping construction, VGO deep-cut operation should be attempted by increasing the vacuum distillation tower inlet temperature (modification cost not required). Then, packing replacement, cooling water temperature reduction by introducing a chiller, and installation of a vacuum pump should be studied.

Although collected data is insufficient and the modification effect cannot be quantitatively calculated, the cooling water piping (10-inch diameter, about 30 m length) modification cost in the above item c. is approximately 3 million yen, which can be repaid in a very short period because of both aspects of energy conservation and production increase effect. As explained below, the current cooling water system (CTW) has little allowance. However, examination including CTW balance review is required.

At the initial stage of this study, possibility of insufficient cooling water supply to the coolers at a height of 10 meters above the ground was checked to find the cause of the insufficient cooling capability of the overhead coolers. As a result of calculating pressure drop ΔP of the cooling water piping (10 inches diameter), there was no problem in the cooling water flow rate.

(6) Steam control

In these plants, there are many steam leak locations from the steam trap and steam piping, and also many steam tracers not heat-insulated were found. Also, there are many leaks of liquid from the gland seal of pumps and valves.

To control steam leak, when a person finds a steam leak, he/she should immediately record it in the notebook. Based on the record, the action such as steam trap replacement should be taken as early as possible. If no action can be taken immediately, a maintenance list and necessary components should be prepared so that a proper action can be taken upon regular maintenance. In these plants, disk type steam traps are mainly used for steam tracers, so disk maintenance is one of the important self-management activity items.

In these plants, the Energy Control Committee is supposed to take the lead in listing of steam leak locations from the South Plant, so the result is expected.

(7) Heat insulation

a. Heat insulation control

In these plants, peeling of the heat insulation material from the steam tracer around the piping or valve was often found. As a basic of the energy conservation activities, the broken heat insulation material should be immediately repaired and a thicker material should be used if necessary.

The heat insulation material is often removed for maintenance around valves. Many years ago, peeling of the steam tracer heat insulation material around valves was often seen in Japan. Recently, a jacket type heat insulation material allowing removal and re-setting is used.

In these plants, energy conservation can be attempted by using a heat insulation material that can be removed and re-set.

To detect the faulty heat insulation, an infrared thermal video system such as AVIO used in this investigation is effective.

b. Steam piping valve heat insulation

Around the boilers in the South Plant, heat insulation is not applied to the steam pipe valve. Since the surface area per unit length of the valve is larger than that of a straight pipe with the same size, heat insulation should be applied like the straight pipe. Table 2.23 lists the equivalent lengths of various valves corresponding to the straight pipe.

Table 2.23 Equivalent Length of Valves

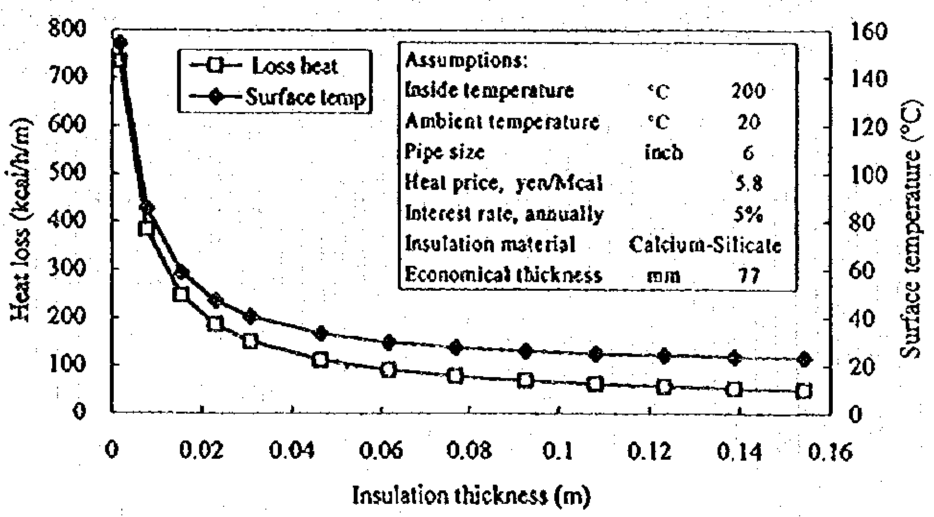
Nominal Size (inch)	Pipe Diameter (mm)	Equivalent Length (m) F-G 10 k	F-G 20 k	F-S 10 k	F-S 20 k
15	21.7	1.15	1.24	1.12	1.29
20	27.2	1.06		0.98	1.13
25	34.0	1.22	1.21	1.15	1.32
40	48.6	1.11	1.20	1.31	1.23
50	60.5	1.11	1.28	1.22	1.53
65	76.3	1.23	1.50	1.16	
80	89.1	1.25	1.56	1.31	1.63
100	114.3	1.27	1.58	1.20	1.50
125	139.8	1.40		1.27	
150	165.2	1.50	1.78	1.35	1.92
200	216.3	1.68	1.87	1.52	

Symbols: F-G Flange ball valve
F-S Flange sluice valve
10 k 10 kg/cm² rating
20 k 20 kg/cm² rating

As shown in this table, a ball valve with an 8-inch pipe diameter has a heat emission area equivalent to approximately 2-meter piping.

When heat insulation is applied to the steam pipe, heat emission sharply reduces according to the thickness of the heat insulation material. Figure 2.15 shows an example.

Figure 2.15 Steam Pipe Heat Insulation (an example)



Many various valves are installed in these plants and the number of the valves is unknown. The sizes and number of valves are assumed to calculate the valve heat insulation effect. Table 2.24 shows the calculation result.

Table 2.24 Heat Economy by Steam Valve Heat Insulation

Size (inch)	No. of Valves	Equivalent Length (m)	Heat Loss (kcal/h·m) Bare Tube	Insulated Tube	Advantage	Total Advantage (kcal/h)
1	700	1.2	198.4	41.9	156.5	125,979
2	700	1.2	353.1	52.7	300.4	256,510
4	560	1.2	667.0	71.1	595.9	400,476
6	350	1.4	964.1	86.8	877.2	414,492
8	210	1.5	1262.3	101.6	1160.7	370,508
12	70	1.7	1858.7	130.3	1728.4	208,103
Total (Norht Plant)						1,776,068

Valves are assumed to be sluice type of 10 kg/cm^2 rating.

When heat insulation is applied to the steam valves, heat emission is reduced to approximatecy 10 %.

The total value is indicated in calories. If the value is converted into fuel assuming that the boiler efficiency is 80 % and operation rate in a year is 93 %, it is 16,278 Gcal/year (= 1,776.068 kcal/h × 0.9 × 8,760 × 0.93/0.8).

During the hearing, the plant people asked us "how should economy of insulation be considered?". We answered that the energy conservation "fuel reduction" amount derived from heat insulation enhancement "reduction in heat emission loss" should be compared with investment such as the heat insulation material cost and construction labor cost.

(8) Operation improvement by chemical injection

In oil refining processes in Japan, the following chemicals (several ppm to tens of ppm in any case) are added for operation improvement:

a. Improvement of separation in the desalter

In this refinery, the salt content in crude oil is high (salt 20 lb/1,000 B).

To maintain the salt removal efficiency, the crude oil desalter section at the upstream of the atmospheric distillation process desalts in two stages. However, when the crude oil tank is changed, the oil-water separation effect reduces and the desalter operation varies.

If the desalter performance degrades, reduction in the heat transfer efficiency or corrosion may occur due to salt precipitation inside the heat exchanger in the downstream process or inside the heating pipe of the heating furnace.

In Japan, oil-water separation troubles in the desalter often occurs along with importing of heavier crude oil. If the crude oil tank has a sufficient retention time, the crude oil is settled to separate and remove salt for as many days as possible. Then, an emulsion breaker such as demulsifier is mixed into the washing water to be injected into the desalter to increase the separation efficiency.

In this refinery, chemicals such as emulsion breaker are currently not injected into the desalter cleansing water. For operation efficiency improvement, a case study including the chemical cost should be implemented.

b. Measures against fouling of the reformer stabilizer overhead condenser

In this study the plant people raised the problems of calcium carbonate ($CaCO_3$) fouling in the reformer stabilizer overhead condenser and subsequent heat transfer pipe corrosion.

The following two measures are considered against calcium carbonate fouling in Japan:

① Addition of a chemical to prevent calcium carbonate adhesion

Domestic and foreign manufacturers have developed various anti-foulants. By collecting information from several manufacturers, an anti-foulant is selected and injected as required.

② Installation of the deionized water circulation system

Presently, the cooling water is circulated by the CTW system using industrial water. Changing it into the steam condensate (pure water) circulation system allows the removal of calcium carbonate itself and the complete prevention of fouling. However, since the equipment cost is high, this case is an energy conservation item in step III.

In addition, fouling prevention by increasing the cooling water flow speed in the cooler can be considered.

However, the current CTW system (total designed circulation: 8,090 tons/h) has an insufficient capability in terms of temperature balance as shown below; therefore it is not a practical action.

Designed supply temperature: 80 °F ≒ 26.7 °C
Actual supply temperature: 95 °F ≒ 35.0 °C

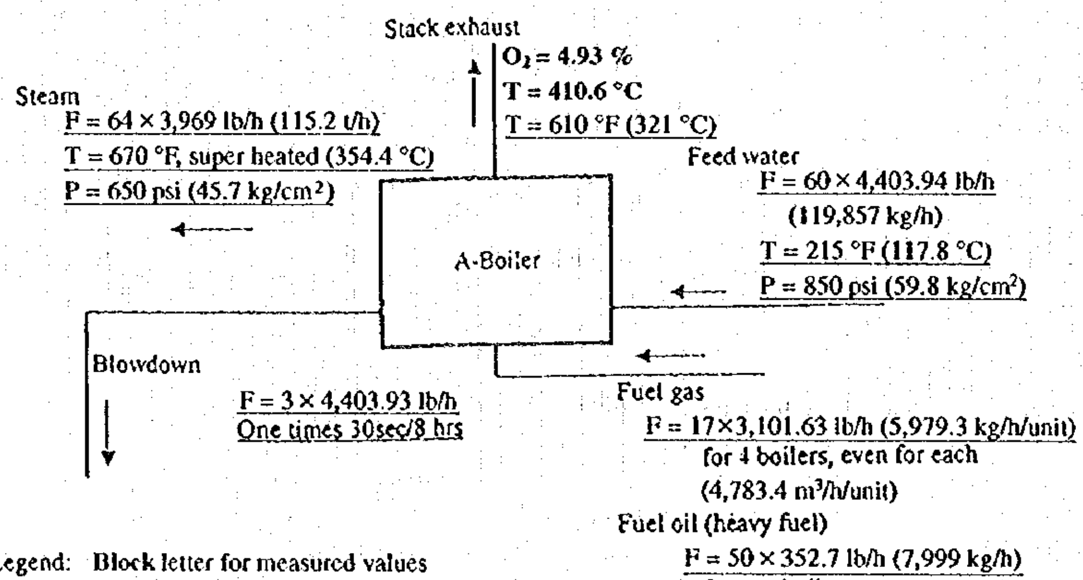
Since there is no detailed data available, the cause for this supply temperature rise is unknown. Anyway, the current CTW system does not have a sufficient capability.

For corrosion of the cooling pipe on the overhead condenser, corrosion may be accelerated by partial battery generation caused by calcium carbonate fouling. Therefore, calcium carbonate fouling should first be prevented by chemical injection. Then, pipe material change into anti-corrosion alloy should be examined.

(9) Boiler

A power plant is provided in each of the North and South Plants for power generation and process steam supply. Among them, the flue duct exhaust gas of boiler A in the South Plant was measured. Figure 2.16 shows the measurement result and meter indication in the control room.

Figure 2.16 Measured Data for Boiler A/South Plant



Legend: Block letter for measured values
Underlined letter for control room reading

T= Temperature
P= Pressure
F= Flow rate

For this boiler, the exhaust gas temperature is very high (measured value: 410 °C) and meter indication in the control room is 321 °C. Therefore, heat loss caused by exhaust gas is large, corresponding to 18 % of fuel heat as shown in Table 2.25 according to calculation.

Table 2.25 Exhaust Gas Heat Loss (South Plant)

O_2 dry	Exhaust Gas Temperature		
	200 °C	300 °C	400 °C
0.0 %	6.4 %	10.3 %	14.3 %
2.5 %	7.1 %	11.4 %	15.9 %
5.0 %	8.1 %	13.0 %	18.0 %
7.5 %	9.4 %	15.1 %	20.9 %
10.0 %	11.3 %	18.1 %	25.1 %

Fuel gas composition (Common in south plant)	
H_2	–
C_1	55.3 %
C_2	8.2 %
C_3	17.9 %
i-C_4	6.3 %
n-C_4	5.5 %
i-C_5	1.0 %
CO	5.8 %
CO_2	Free
H_2S	15 ppm

As shown in the table, a large amount of heat is lost as the exhaust gas sensible heat. The heat transfer surface should be cleaned to improve heat absorption by the boiler. Furthermore, if necessary, an air preheater should be installed to recover the exhaust gas heat efficiently. If the fuel air is preheated by the exhaust gas heat, the fuel saving rate 8 % (if air is preheated to 200 °C) as shown in Table 2.26 according to calculation.

Table 2.26 Fuel Economy by Air Preheating

Exhaust gas temperature before preheating = 410.6

O_2 dry	Air Preheating Temperature		
	30 °C	200 °C	250 °C
0.0 %	0.0 %	6.2 %	7.9 %
2.5 %	0.0 %	7.1 %	9.0 %
5.0 %	0.0 %	8.2 %	10.4 %
7.5 %	0.0 %	9.8 %	12.4 %
10.0 %	0.0 %	12.3 %	15.4 %

If the same method can be applied to the North Plant, fuel saving in the North Plant is as follows:

Total steam volume in the North and South Plant:	720 tons/h
Steam volume in the North Plant:	2,607 × 10^3 tons/year
Steam calories (354.4 °C, 45.7 kg/cm^2):	739.2 kcal/kg

If the boiler efficiency is 80 %, the saved volume is 21,177 kL/y (= 2,607 × 10^3 × 739.2 × 10^3 / (9,100 × 10^3) × 0.08).

(10) Turbine

In response to the request of the plant, the Japanese study team measured the turbine efficiency on June 17. No main steam flow meter for the steam turbine was available, although the main steam pressure of the boiler and the temperature were indicated on the chart in the boiler room. Therefore, we temporarily installed an ultrasonic flow meter to measure the turbine condensate flow rate. This turbine is operated without extraction steam, and the boiler is not equipped with any feedwater preheater and uses the steam from the external line (steam from the other line than the power plant line) for the deaerator and the ejector.

Table 2.27 shows the specifications of the boiler, the turbine and the generator, and Figure 2.17 and Table 2.28 show the results of measurement. The graph line representing the relationship between the power generation output and the steam flow rate crosses at the point (0, 0), suggesting that the condensate flow rate might have probably been measured to be about 2 t/h smaller. Hence, by obtaining an approximate formula to correct the condensate volume by 2 t/h, steam consumption rate at 5 MW will be 4.76 kg/kWh. These equipment still keep the same capacity as when they were newly installed, showing that they are well maintained.

In addition, steam flow rate can be estimated based on the pressure after the Curtis stage unless the turbine blade becomes dirty. Thus, it is recommended that the characteristic curve for the steam flow rate estimated from the pressure after the Curtis stage should be obtained from the manufacturer to be utilized for controlling the performance.

The isentropic efficiency and the thermal efficiency of a turbine generator at 5 MW of power generation are as shown below.

(Isentropic efficiency of a turbine and a power generator)

$$\eta_{ctg} = \frac{860P}{G\,(i_1 - i_2)} = \frac{860 \times 5{,}000}{23{,}800\,(737.7 - 482.5)} = 70.9\ \%$$

(Thermal efficiency of a turbine and a power generator)

$$\eta = \frac{860P}{G(i_1 - i_0)} = \frac{860 \times 5{,}000}{23{,}800\,(737.7 - 38.0)} = 25.8\ \%$$

where i_1: enthalpy of steam at 45.5 ata and 352.2 °C (kcal/kg)

i_2: enthalpy of steam when the steam under the same conditions as i_1 is adiabatically expanded to 50 mmHg (0.068 ata) (kcal/kg)

i_0: enthalpy of feedwater at the time mentioned above

Table 2.27 Specifications of Boiler, Turbine and Generator

Boiler	Steam condition	650 psig, 750 °F
	Feed water temperature	220 °F
	Boiler efficiency	79 %
Turbine	Steam condition	625 psig, 700 °F (43.9 kg/cm²g x 371.1 °C)
	Vacuum	4.5 inch Hg (114.3 mmHG abs)
	Rated output	7,200 kW
	Speed	6,500 rpm
	Type	Impulse turbine
	Stage	22 stage (according to drawing)
	Steam consumption	10.4 lbs/kWh (4.722 kg/kWh)
	Manufacture	Siemens (1973)
Generator	Capacity	9,000 kVA (7,200 kW)
	Power factor	0.8
	Voltage	6,300 V
	Current	825 A
	Insulation	IPR 44

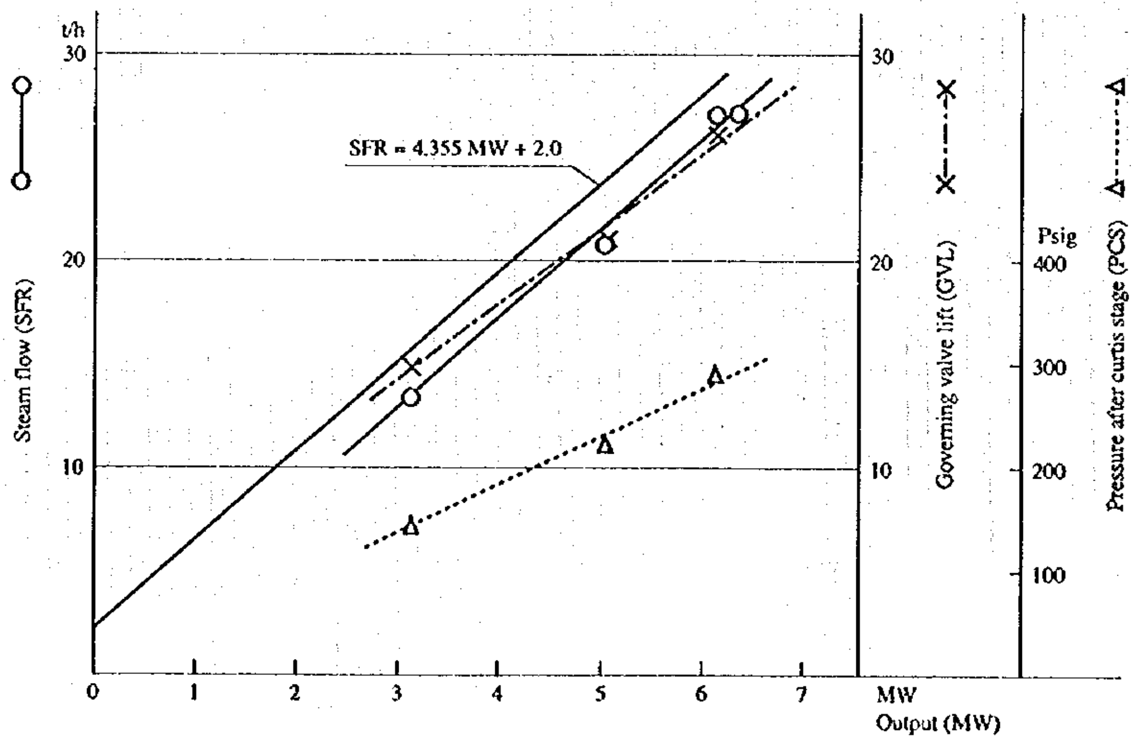
Table 2.28 Test Record

Date: June 19, 1996

		9:32	9:44	10:15	10:39	10:50	Meam
Boiler steam pressure	psig	2A/2C 645/630		645/620			635
Boiler steam temperature	°F	655/672	666	663/672			666
Turbine pressure after MSV	psig	590	592	586	582	582	588
Turbine temperature after MSV	°F	640				640	640
Stam pressure after curtis stage	psig	218	217	285		140	–
Vacuum	inch	–24.0	–24.0	–23.5		–24.5	24.0
Governing valve lift			21.0	26.0	26.5	15.0	–
Generator output	MW	5.0	5.0	6.1	6.3	3.1	–
Condensate flow rate t/h		20.8	20.4	27.0	27.1	13.3	–
Condensate temperature after ejector	°F		103	103		103	103
Turbine exhaust temperature	°F		108	115			
Cooling water temperature	°F		84/104				

Note: Estimated atmospheric pressure 659 mmHG based on 1,191 m above sea
Estimated condensor vacuum 50 mmHG

Figure 2.17 Turbine Performance Record



(11) Electrical system

This refinery is under such restrictions as in-house power generation. From the viewpoint of safe and stable operation, energy conservation has limitations. This problem is basically attributable to the Iran's unstable power system. Some other plants are operating private power generation under low utilization factor of the power generator to secure safe and stable operation.

To solve this problem, the Iran's power system should be enhanced. If its reliability increases, cases of parallel operation with the system will increase. If the power factor discount charge is applied as in Japan, power factor improvement will be proceeded.

Although the power factor (83 %) in the entire North Plant is not satisfactory, reduction in resistance loss alone will not be enough as a merit of condenser installation. When parallel operation with the external power system becomes reliable in future, the problem should be solved along with the said problem of equipment utilization rate.

The reactor at the connection point with the South Plant may be necessary for stable operation. But it is not favorable because loss arises and a voltage difference occurs if current flows between the North and South Plants. Therefore, it is desirable for both South and North Plants to operate the power generating facilities so as to keep these facilities and the load in good balance.

For the load balance between power generators 2B and 2C, balancing is recognized on active power but slight unbalancing is found on reactive power. If reactive power is also balanced, loss from the power generators will be reduced and power distribution loss from load bus lines A and B will reduce. Therefore, balanced operation is required.

(12) Pump

Table 2.29 shows the current measurement result for large motors relating to the distillation column and visbreaker. As shown in the table, most large motors are used for pumps. During this measurement, connection to the voltage terminal was difficult, so current only was measured. In this table, the current ratio means the ratio of the measured current value to the rated current.

Table 2.29 Large Motor Current Measurement Result

Motor No.	Motor Capacity (HP)	Rated Current (A)	Measured Values of Current (A)	Current Ratio	Remarks
2PM153	250	22.5	20.7	0.92	High voltage
2PM154	400	35	25.5	0.73	High voltage
2PM108	250	22.5	17.0	0.76	High voltage
2PM157	300	27	24.0	0.89	High voltage
2PM105	300	27	20.0	0.74	High voltage
2PM311	250	22.5	21.5	0.96	High voltage
2PM302	350	31.5	8.4	0.27	High voltage
2PM301	350	31.5	21.0	0.67	High voltage
2PM106	150	190	190	1	Low voltage
2PM107	125	160	107	0.67	Low voltage
2PM156	75	95	87	0.92	Low voltage
2EM162	25	35	29	0.83	Low voltage
2EM109B	30	40	31	0.78	Low voltage
2EM109D	30	40	29	0.73	Low voltage

Note: PM in the motor No. stands for pump, while EM stands for fan.

a. Motor replacement

According to the current ratio, the 2PM302 350 HP motor seems to involve a problem. Power at this point is $350 \times 0.735 \times 0.27/2 = 35$ kW according to calculation using the current ratio. If this motor is replaced with a 37 kW (50 HP) motor, 90 % can be expected as its efficiency. Efficiency of the 350 HP motor is assumed to be approximately 85 % for this load. Therefore, the axial power at this point is $35 \times 0.85 = 29.8$ kW.

On the other hand, input of the 37 kW motor is $29.8/0.9 = 33.1$ kW; therefore power saving as a result of motor replacement is $(35 - 33.1) \times 8{,}000 = 15{,}200$ kWh (on the assumption of 8,000 hour operation in a year) and power cost reduction is $15{,}200 \times 10$ yen/kWh = 152,000 yen. The required investment is 740,000 yen if the motor installation cost is 20,000 yen/kW. Therefore, the simple investment recovery period is 5 years.

b. Impeller cutting

In this study, a large pump with high power consumption was selected and the opening degree of the CV (control valve) in the delivery line was checked to examine the possibility of energy conservation (power saving) by cutting the impeller of the pump. Figure 2.18 shows the flow outline around 2P-105 A&B and measured data. Table 2.30 shows the design base of the related pumps and actual opening degree of CV.

Figure 2.18 Electric Power Conservation by CV Opening

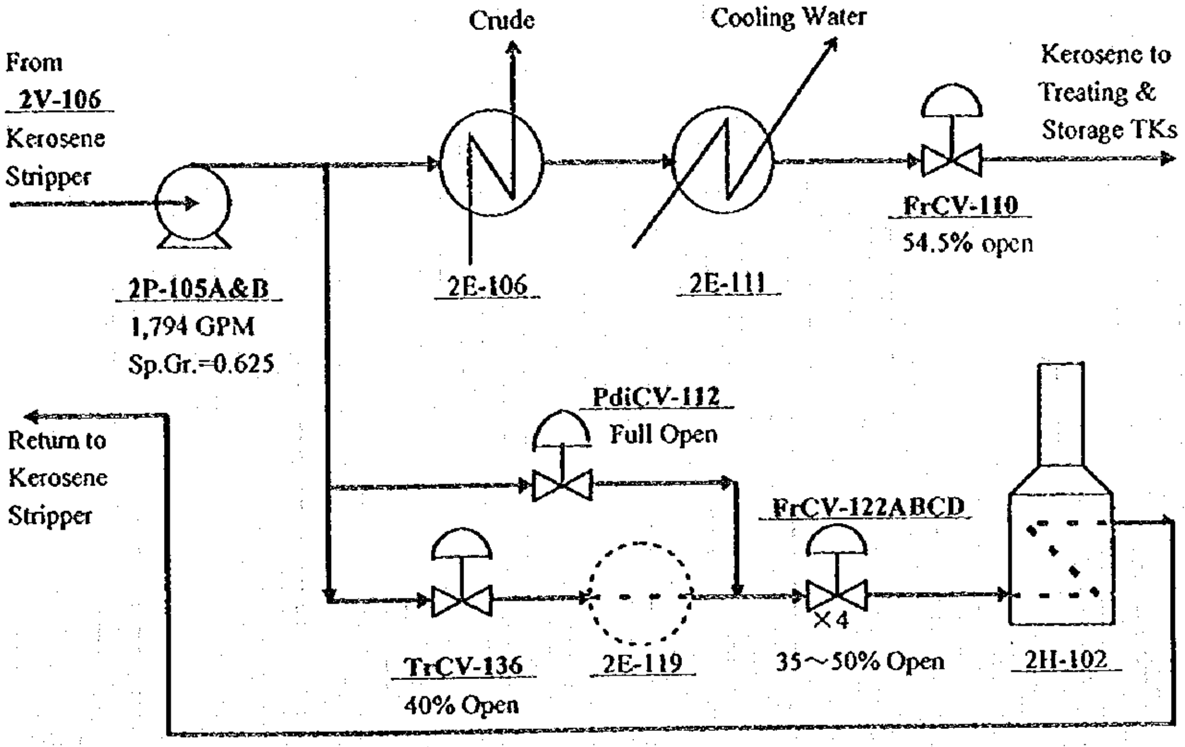


Table 2.30 Design Base & Actual Data of Pump & Control Valve

Date	1996	Jun. 18		Jun. 18	
Time Checked		@10:20		@10:25	
Tag No.		2P-105 A&B		2P-107 A&B	
Service		Kerosene product		Light diesel to storage	
Specific gravity	-	0.625			
Flow rate	gpm	1,794		621	
Electric power required	hyHP/brHP	195/238			
Operating temperature	°F	247			
Differential pressure	psi	11.3			
Control valve		FrCV-110	PdiCV-112	FrCV-108	FrCV-125
CV & connecting line	inch (Dia.)	6"-3"-6"	TrCV-136	6"-4"-6"	
CV value	-	Req 58/Eq % 116		Req 91/Sel 196	
Differential pressure	psi	Ratd 5.0/Max 17.3	Full open	Ratd 1.5/Max 12.7	
Actual CV opening	%	54.5	40	17	63
Remark		Impellers should be cut.		Impellers should be cut.	

Date	1996	Jun. 18		Jun. 18	
Time Checked		@14:40		@14:45	
Tag No.		2P-153 A&B		2P-154 A&B	
Service		Heavy diesel sidestream		Heavy vacuum gas oil	
Specific gravity	-	0.757		0.73	
Flow rate	gpm	1,400		2,134	
Electric power required	hyHP/brHP	182.5/225		242/301	
Operating temperature	°F	129		257	
Differential pressure	psi	14.1		11.5	
Control valve		FrCV-171	PdiCV-177	FrCV-168	FrCV-173
CV & connecting line	inch (Dia.)	6"-4"-6"	6"-4"-6"	8"-4"-8"	6"-6"-6"
CV value	-	Req 131/Sel 196	Req 56/Sel 116	Req 127/Sel 183	Req 449/Sel 305
Differential pressure	psi	Ratd 6.7/Max 15.3	Ratd 1.8/Max 13.4	Ratd 1.1/Max 10.5	Ratd 3.2/Max 12.7
Actual CV opening	%	100	35	100	32
Remark		Impellers should be cut with CV size-up.		Impellers should be cut with CV size-up.	

Note: hyHP/brHP: Hydraulic/Breaked HP
Req : Required
Sel : Selected
Eq : Equal
Ratd : Rated

The Cv formula as a basis for this examination is as follows:

$$Q = Cv(\Delta P/(Sp.Gr.))^{1/2}$$

where,

Q:	Fluid flow rate in CV, gpm
Cv:	CV native value determined by the opening degree
ΔP:	Pressure drop caused by CV, psi
(Sp.Gr.):	Specific gravity of the fluid in CV

In the case of maximum load during normal operation, the CV consumes pressure drop larger than Δp required for control for a system with a small opening degree of the CV. Therefore, for the portion of pressure drop more than required, power can be saved by cutting the impeller of the pump until the delivery pressure allows the CV opening degree to be in the proper level (normally 80 %). As a result of study, it was found that the impeller could be cut for 2P-105 A or B and 2P-107 A or B. Power conservation shown in Table 2.31 is expected to be achieved with these two pumps.

Table 2.31 Electric Power Reduction by Pump Impeller Cutting

Tag No.	Capacity kL/h	Motor Capacity HP	Estimated Reduction MWh/y	Modification Cost k¥	Simple Payout Period y
2P-105 A or B	407	300	535	2,000	0.23
2P-107 A or B	141	125	141	1,000	0.44
2P-153 A or B	304	250	123	3,100	1.58
2P-154 A or B	485	400	100	3,900	2.44

If power conservation cannot be attempted because the current CV size is small and the required pressure drop is large, power conservation by impeller cutting is allowed after increase of the CV size (CV replacement). For 2P-153 A or B and 2P-154 A or B, the opening degree of some CVs of the delivery piping is close to "fully open". If the size of the CV is increased, impeller cutting is allowed. Table 2.31 shows the modification cost and power conservation including CV replacement of these two pumps.

As described above, this examination is based on the assumption that the load is normally close to the maximum level in this refinery. In both cases, impeller cutting is only applied to the pump that is normally in service among pumps A and B. Impeller cutting should not be applied to the spare pump because even a large increase in operation load can be coped with.

(13) Lighting

As indicated in the preliminary study (October, 1995), outdoor lighting (mercury lamp and sodium lamp) in the daytime is still recognized. According to visual check, there are approximately 100 lamps in both plants. Turning off the unnecessary lamps is the first step to energy conservation. Although circuits may have to be grouped to a certain degree, this action does not require cost and should be implemented immediately.

If one hundred 250 W sodium lamps can be turned off in the daytime, 365 days × 10 hours × 100 lamps × 0.25 kW = 91,250 kWh can be saved in a year.

(14) Others

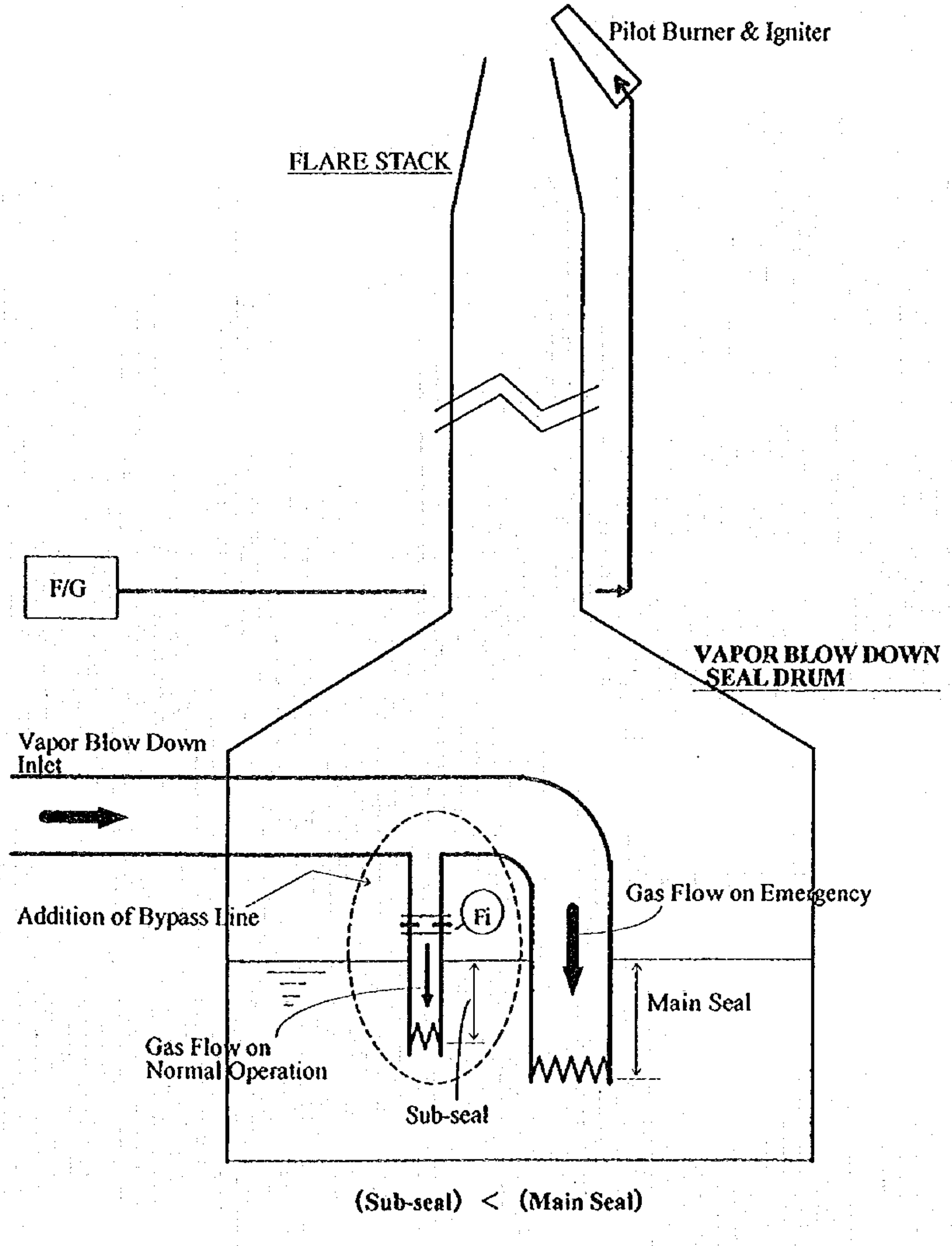
a. Flare system improvement

The operation status of an oil refinery is concentrated into the flame size in the flare system. In other words, if the operation of some equipment in the refinery changes, the flare flame size increases. If large-loss operation is performed, the flare flame size also increases. In the refinery, the flare size (gas volume) should be daily checked, controlled and reduced as an energy conservation promotion means.

During the hearing, the plant people requested us to measure the flow rate in the current flare system. However, the flare piping normally has a "positive pressure" of hundreds of mmH_2O and the flare gas coming out during installation of the measuring equipment normally contains highly poisonous H_2S gas. Also, because of a time limitation, we could not measure the flare gas flow speed.

As an alternative, Figure 2.19 shows a method of installing a bypass piping for measuring the flare gas flow rate during normal operation. Since the diameter of the main seal piping is large for emergency cases, the gas flow speed is normally low and the value cannot be measured. As shown in Figure 2.19, if a small diameter bypass piping with a seal height smaller than that of the main seal is installed, gas flows to the bypass piping during normal operation. Therefore, the flow speed in the bypass piping increases and the flare gas flow rate can always be measured.

Figure 2.19 Improvement of Flare Loss Control



(Sub-seal) < (Main Seal)

b. Effective utilization of sour water

In the refinery, a large amount of sour water containing hydrogen sulfide is generated as a result of introducing the process steam such as steam injection. Normally, the sour water is processed by steam stripping and reused or discharged. Depending on the use, sour water not processed can be used for energy conservation (e.g. diluting water for alkali injection into crude oil, desalter washing water, etc.).

c. Safety measures

In plants in Japan, a flame arrestor net is attached to the muffler of the vehicle that entered the plant to prevent ignition of gas accumulated in the manhole under the load.

Therefore, we proposed a method of attaching a net to this refinery.

(15) Conclusion

Table 2.32 summarizes improvement items described above.

Table 2.32 Summary of Proposals

(Japanese Yen base)

Item	Expected Saving: Fuel kL/y	Fuel Million yen/y	Fuel %	Electricity MWh/y	Electricity Million yen/y	Electricity %	Total Million yen/y	Investment Million yen	Payback Period Year
Improvement of the reheating furnace inside refractory	538*1	9.1	0.1*6	–			9.1	20	2.2
Improvement of the reheating furnace air ratio	16,983*2	288.7	2.4*7	–			288.7	90	0.3
Pre-heating of the reheating furnace air	27,053*3	459.9	3.8*8	–			459.9	1,795	3.9
Enhancement of heat recovery from the cooler	1,781*4	30.3	0.2*9	–			30.3	62	2.0
Heat insulation of the steam pipe valves	1,789*5	30.4	0.2*10	–			30.4	115	4.1
Pre-heating of the boiler combustion air	21,177	360.0	3.0*11	–			360.0	1,649	4.6
Replacement of pump motors	–			15.2	0.2	0.02*12	0.15	0.7	5.0
Pump impeller cutting	–			899	9.0	1.0*13	9.0	3	0.3
Turning off unnecessary lights	–			91.3	0.9	0.1*14	0.9	0	0
Total	69,321	1,178.4	9.7	1,005.5	10.1	1.1	1,188.45	3,734.7	3.2

(Iran Rial base)

Item	Expected Saving: Fuel F.oil kL/y	Fuel Million Rial/y	Fuel %	Electricity MWh/y	Electricity Million Rial/y	Electricity %	Total Million Rial/y	Investment Million Rial	Payback Period Year
Improvement of the reheating furnace inside refractory	538*1	40	0.1*6	–			40	350	8.8
Improvement of the reheating furnace air ratio	16,983*2	1,274	2.4*7	–			1,274	1,575	1.2
Pre-heating of the reheating furnace air	(27,053)*3	(2,029)	(3.8)*8	–			(2,029)	(31,413)	(15.5)
Enhancement of heat recovery from the cooler	1,781*4	134	0.2*9	–			134	1,085	8.1
Heat insulation of the steam pipe valves	(1,789)*5	(134)	(0.2)*10	–			(134)	(2,013)	(15.0)
Pre-heating of the boiler combustion air	(21,177)	(1,588)	(3.0)*11	–			(1,588)	(28,858)	(18.2)
Replacement of pump motors	–			15	2	0.02*12	2	12	6.0
Pump impeller cutting	–			899	90	1.0*13	90	53	0.6
Turning off unnecessary lights	–			91	9	0.1*14	9	0	0
Total	19,302	1,448	2.7	1,005	101	1.1	1,549	3,075	2.0

*1 $4{,}893 \times 10^3$ Mcal/y/$9{,}100 \times 10^3$ kcal/L = 538 kL/y
*2 $154{,}546 \times 10^6$ Mcal/y/$9{,}100 \times 10^3$ kcal/L = 16,983 kL/y
*3 $246{,}179 \times 10^6$ Mcal/y/$9{,}100 \times 10^3$ kcal/L = 27,053 kL/y
*4 2×10^6 kcal/h $\times 24 \times 365 \times 0.925/9{,}100$ kcal/L = 1,781 kL/y
*5 $16{,}278 \times 10^6$ Mcal/y/$9{,}100 \times 10^3$ kcal/L = 1,789 kL/y
*6 $538 / (1{,}433 \times 10^3 / 2) \times 100 = 0.1$ %
*7 $16{,}983 / (1{,}433 \times 10^3 / 2) \times 100 = 2.4$ %
*8 $27{,}053 / (1{,}433 \times 10^3 / 2) \times 100 = 3.8$ %
*9 $1{,}781 / (1{,}433 \times 10^3 / 2) \times 100 = 0.2$ %
*10 $1{,}789 / (1{,}433 \times 10^3 / 2) \times 100 = 0.2$ %
*11 $21{,}177 / (1{,}433 \times 10^3 / 2) \times 100 = 3.0$ %
*12 $15.2 / 188{,}650 / 2 \times 100 = 0.02$ %
*13 $899 / 188{,}650 / 2 \times 100 = 1.0$ %
*14 $91.3 / 188{,}650 / 2 \times 100 = 0.1$ %

Energy price in Japan:
Fuel price: 17,000 yen/kL
Electricity price: 10 yen/kWh

Energy price on Iran Rial base:
Fuel oil: 75 Rial/L
Electricity: 100 Rial/kWh

Exchange rate: 1,750 Rial = 1 US Dollar = 100 Japanese Yen
Calorific value of fuel: Oil: 9,000 kcal/kL

Investment cost is based on that in Japan.