(iii) Flare Stack

A flare stack is located at off-site area with sufficient distance from the oil gas separation facilities for safety. The gas separated at each separator was transmitted, after separating oil mist in the scrubbers to the flare stack and was burnt there.

(iv) Building

There exist one warehouse and one electric power generator house.

5-2 Prospect of Natural Gas Production

The discussion in this section has the aim of estimating the delivery rate, pressure, temperature, and composition of the gas and condensate at the outlet of the GOSP. For this purpose, discussed in the following are the production performance of the natural gas reservoir, the entire production system performance and the operation conditions of the GOSP.

The production performance of the natural gas reservoir is the principal factor which exerts a dominant influence on the delivery conditions. However, as stated at the beginning of this chapter, the availability of the reservoir data in this study is so limited that the discussion had to be developed, exercising best engineering judgment in many aspects.

5-2-1 Behavior of Natural Gas Reservoir

In the gas utilization program like this project, it is essential to grasp the reservoir behavior to some extent. Discussed in the following are considerations on the characteristics, fluid composition, and production performance of the reservoir:

(1) Reservoir Characteristics

The depth, temperature, and pressure of the reservoir are the basic characteristics for estimating its behavior. Since given data for the reservoir are rather limited, they are assumed or estimated as follows:

(a) Reservoir Depth

In this project, the natural gas is intended to be produced from the ten layers, Zones M/N/O3/P1/P2/P3/P4/Q/R/S, in the Talang Akar sandstone formation. Judging from the general geological information, the depth of the entire

Talang Akar sandstone formation in which the target ten zones exist is considered to range from 1,600 to 1,900 m at maximum.

As for the data on the depth, only the following test intervals (depth interval of the portions perforated as a target zone) are given in the production test and gas analysis report for SNT-3 (see Table 5-4). Since Zones M to S are considered to be named in the order of depth (M/N/O3/P1/P2/P3/P4/Q/R/S), both P4 and R, given below, are considered to be located at deeper positions among them.

Tested Zone	Test Interval (m)
P4	1,784.5 - 1,787.5
R	1,784.5 - 1,787.0

On the basis of the above, the natural gas reservoir was regarded as a reservoir composed of the above ten zones, and the depth of the reservoir was assumed as follows:

Datum Depth	:	1,770 m
Range	:	Approx. 1,700 - 1,800 m

Table 5-4 Gas Analysis Data in Production Test Report

Well No.		SNT-3	SNT-3
Layer or Zone		P4	R
Producting/Tested Interval (m)		1,784.5 - 1,787.5	1,784.5 - 1,787.0
Choke Bean Size (mm)		7	11
Date of Sampling		1-5-1974	3-5-1974
Density @ 28°C (g/l)	SMS1316*	0.92	0.91
Absorption & Combustion Analysis (% Vol.)	SMS1322*		
CO_2		7.5	3.6
O_2		0.0	0.0
N_2		0.0	0.0
CnH2n+2		92.5	96.4
		100.0	100.0
Hydrocarbon Analysis (% Vol.)	GLC*		
C1		59.1	57.8
C2		15.1	14.8
C3		15.4	13.9
iC4		3.3	3.8
nC4		4.4	5.9
iC5		1.7	2.3
nC5		1.0	1.5
C6+		0.0	0.0
		100.0	100.0
Specific Gravity (air=1)		0.74	0.75

(Data Source by PERTAMINA)

Note: * denotes the method of the analysis.

(b) Reservoir Temperature

The temperature of the reservoir was estimated according to the average ambient temperature at the City of Jambi and the geo-thermal gradient data in the Sengeti district. The average ambient temperature was assumed to be 26.7°C (80°F) according to the measured data for the past eight years. As for the geo-thermal gradient, 4.93°C/100 m (2.7°F/100ft) was obtained as the data on SNT-2 from the publication of Indonesian Petroleum Association, "Geo-thermal Gradient Map of Indonesia."

Based on these data, the reservoir temperature was estimated by applying the following equation:

Reservoir Temperature = Surface Temperature
+ Geo-thermal Gradient x Depth

Therefore,

Average Reservoir Temperature :	113°C (235°F)
Temperature Range of Reservoir:	approx. 110 - 115°C (230 - 239°F)

(c) Reservoir Pressure

The reservoir pressure was estimated based on the assumed datum depth and the measured static wellhead pressures given in Table 5-5. As the data in Table 5-5 were obtained in July, 1985, the reservoir pressure as estimated below can be regarded as current pressure.

Table 5-5 shows both the tubing and casing pressures. There is no data on the well completion, but it is expected that a production packer is installed at the bottom-hole, judging from the standard practice for completion in PERTAMINA. If this is true, the tubing side and the casing side are isolated from each other, and there should have been no continuity of pressure between them. However, the measured tubing and casing pressures for all the production wells except SNT-10 are equal to each other. This might be attributed to the possibility of tubing leakage in the wells which might have occurred after the suspension of production.

The static wellhead pressure (tubing-top pressure) varies with the wells. Among them, SNT-4, SNT-9 and SNT-17 have extremely low pressures compared with the others. This is considered to be due to some trouble in the wellbore. The wellhead pressures of the wells other than the above three range from 125 to 145 kg/cm^2G. Many of them are around 130 kg/cm^2G, and their average works out at 131 kg/cm^2G. Estimation of the reservoir pressure was based on this average static wellhead pressure.

The static wellhead pressure is the tubing-top pressure when the well is shut-in. When the well is shut-in, the bottom-hole pressure is equal to the static reservoir pressure, and the wellhead pressure is lower than two pressures mentioned above by the pressure of the static fluid column. In the case of oil reservoir, the static fluid column consists of two zones, the lower liquid column and the upper gas column. However, as described later, the subject reservoir can be basically regarded as a natural gas reservoir. Therefore, the static bottom-hole pressure (i.e., reservoir pressure) can be estimated by adding the static gas column pressure to the wellhead pressure. It is obtained by applying the following equation (by Pierce and Rawlins):

Table 5-5 Results of Well Static Pressure Test

Well No.	Casing Pressure (kg/cm^2G)	Tubing Pressure (kg/cm^2G)
2	130	130
3	138	138
4	40	40
5	125	125
6	killed	killed
7	128	128
9	84	84
10	15	145
14	125	125
15	130	130
17	42	42
19	130	130

(Data Source by PERTAMINA)

Notes:

1) Test Date : July '85
2) Dry-Holes : Well No.1, 8, 11, 12, 13, 16 and 18.
3) Absolute Open Flow Potential (AOF) = 5 MMSCFD/well.
   Average Gas Deliverability = 2 - 3 MMSCFD/well.

$$P_{ws} = P_{wh} \exp (r_g D/29.26T)$$

where,

P_{ws} : Static bottom-hole pressure

P_{wh} : Static wellhead pressure = 132 kg/cm^{2}A

exp : Base of the natural logarithm

r_g : Specific gravity of the gas (air = 1)
= 1.002 (calculated from the composition)

D : Well depth = 1,770 m

T : Mean temperature in the wellbore = 343°K

The actual reservoir pressure may be a little different from the estimated value, but no further investigation can be done with the available data. Consequently the succeeding discussions are made based on the estimated reservoir pressure value as given below:

Current Reservoir Pressure at the Datum Depth:
158 kg/cm^{2}A (= 2,250 psia)

(2) Type of Reservoir and Fluid Composition

In analyzing the behavior of the reservoir and the production system, the reservoir fluid properties are essential factors. In general, the reservoir fluid is a mixture whose principal components are paraffin hydrocarbon compounds (C_nH_{2n+2}). The fluid may be in the state of either vapor or liquid phase, or in the state of two-phase equilibrium according to temperature and pressure. The vapor-liquid equilibrium state can be analyzed by estimating the equilibrium coefficient of each component. As the methods for estimating the equilibrium coefficient, various equations of state, e.g. Benedict-Webb-Rubin (BWR), Peng-Robinson (PR), Soave-Redlich-Kwong (SRK), etc., are used. For the analysis under high temperature and pressure, such as that of reservoir fluid behavior, the SRK equation is suitable because of its wide range of applicable temperatures and pressures. In the following equilibrium calculations as well as the vapor/liquid enthalpy calculations, the SRK equation is applied.

The reservoir fluid composition and the estimation of the reservoir characteristics according to the composition are discussed below:

(a) Estimation of Reservoir Fluid Composition

As the data on the reservoir fluid composition, the gas analysis data for the ten layers, Zones M to S, are given (see Table 5-6). However, there is some question about interpretation that these data actually represent the reservoir fluid composition itself. The gas contains little amount of hexanes (C6) and heavier components. Obviously, the gas yields little amount of condensate in the stock tank. Other than the data, a sample gas analysis on the production tests for the two zones in SNT-3 is shown in Table 5-4. The compositions are much the same as those in Table 5-6. Contrary to these data, the production history and the production test data in Table 5-1, previously given, indicate a considerable amount of consensate production. To explain the discrepancy, the following three interpretations could be made:

1) The original reservoir was composed of an extremely thin liquid or wet gas (which yields sufficient liquid at the surface) zone and a thick dry gas zone. The former would be made of mainly Zone S, while the latter be made of the other layers. The liquid zone was almost depleted during the past production period, and only the dry gas zone remains at present.

2) During the past production period, the producing layers were different from the ten zones because of the aim of crude oil production. In other words, since only gases existed in the ten zones, no perforation (to conduct the production from the zone) was made.

Table 5-6 Gas Compositions for Ten Zones in Sengeti Field

Component (% vol.)	Reservoir Zone									
	S	R/840	Q/800	P4/790	P3/770	P2/760	P1/750	03/720	N/650	M/640
CO_2	6.3	7.4	7.5	3.6	8.0	7.0	7.0	3.4	5.6	5.7
O_2	0	0	0	0	0	0	0	0	0	0
N_2	0	0	0	0	0	0	0	0	0	0
CnH2n+2	93.7	92.6	92.5	96.4	92.0	93.0	93.0	96.6	94.4	94.3
Hydrocarbon Analysis										
C1	58.7	58.0	60.0	57.8	52.4	58.3	56.8	52.8	49.1	45.7
C2	15.7	15.6	13.8	14.8	17.7	14.3	15.2	15.8	17.5	17.5
C3	14.3	14.6	13.3	13.9	17.3	15.2	15.2	17.2	19.2	21.2
iC4	3.3	3.5	4.1	3.8	3.1	3.8	3.7	3.9	3.3	4.6
nC4	4.4	4.7	6.0	5.9	6.0	5.1	5.1	6.5	5.7	6.8
iC5	1.4	2.2	1.4	2.3	2.3	2.0	2.3	2.6	3.0	2.4
nC5	1.0	1.4	1.4	1.5	1.2	1.3	1.7	1.2	2.2	1.8
C6+	1.2	0	0	0	0	0	0	0	0	0
Sp. Gr. @28°C	0.93	0.96	0.95	0.91	1.01	0.90	0.94	0.95	1.01	1.01

(Data Source by PERTAMINA)

3) The data in Table 5-6 do not show the reservoir fluid composition itself. That is, they are the separated gas analysis obtained from the surface test separation.

In the case of 1) and 2), Table 5-6 shows the reservoir fluid composition itself; however, in the case of 3), it does not and, therefore, the reservoir fluid composition must be estimated in some way.

In general, there are two ways for the reservoir fluid sampling and analysis. In one method, the reservoir fluid obtained by bottom-hole sampling is directly analyzed. In the other method, the gas and liquid samples taken at the surface test separator are analyzed, and then are recombined in calculation at the measured ratio of gas to liquid to obtain the reservoir fluid composition. However, neither Table 5-4 nor Table 5-6 gives the data concerning the condensate production at the time of the gas sampling. There is no proof supporting for the interpretation 3), and the data needed to estimate the reservoir fluid composition under this interpretation are insufficient.

In conclusion, the data in Table 5-6 were interpreted as in either 1) or 2), and the composition given below (mol%) is assumed as the representative one for the whole reservoir. The composition was obtained by taking the simple average of the fluid compositions for the ten zones, since the reserves and the thickness of the individual zones are unknown.

CO_2	C1	C2	C3	i-C4	n-C4	i-C5	n-C5	C6+
6.15	51.57	14.82	15.15	3.48	5.28	2.06	1.38	0.11

Though the properties of the hypothetical component (C6+) are not given, the molecular weight, boiling point and the specific gravity (in the liquid state) must be known for the vapor-liquid equilibrium calculation. The specific gravity of the gas at the temperature of 28°C, shown in Table 5-6, does not agree with that calculated by a mixing rule for the composition; thus, the molecular weight of the hypothetical component cannot be calculated reversely from the specific gravity given. Judging from the composition, the hypothetical component is assumed to have the properties shown below which are almost equal to those of n-heptane:

Molecular weight: 100.2
Boiling point : 98.4°C (209.2°F)
Specific gravity : 0.686 g/cc (42.8 lb/cu-ft)
(in the liquid state)

Further, although water content is not shown in Table 5-6, the reservoir gas is assumed to contain saturated water at given temperature and pressure conditions, as observed in many gas fields.

(b) Type of Reservoir

Reservoirs can be classified into many types; gas, gas condensate, volatile oil, black oil reservoirs, etc. The type of a reservoir can be distinguished according to the relation between the fluid phase behavior and the initial reservoir temperature/pressure. In order to predict the reservoir production performance, the reservoir type was estimated according to the fluid composition assumed in the above. Since the initial reservoir condition is not known, the following discussion is based on the current reservoir condition.

Figure 5-5 shows the pressure-temperature phase diagram for the assumed reservoir fluid. The point C where the boiling point curve A-C and the dew point curve B-C merge is called the critical point. Inside the phase envelope ACB, the fluid is in the two-phase region. Within this area, change of pressure and temperature causes continuous change of the compositions of the vapor and liquid phases, and on the quality lines, X1-C, X2-C, X3-C, and X4-C, liquid molar fraction is constant. The shaded area is called the retrograde region, where the amount of liquid increases as the pressure decreases or the temperature increases, contrary to the normal phase behavior.

The estimated current reservoir pressure, 158 kg/cm^2A (2,250 psia), and temperature, 113°C (235°F) are positioned on the point R. The fluid at this point is in the vapor phase (the super-critical state, to be more specific). The reservoir pressure will be decreased along with the future production, but the temperature will remain constant. The straight line R-D, which indicates the transition, does not cross the two-phase region but is located inside the vapor phase region. Therefore, no retrograde condensation will take place in the reservoir; the reservoir can be classified as a single-phase gas reservoir.

For reference, an example of the temperature and pressure transition while the production fluid flows through the tubing, choke, and flow line to the separator is shown by the broken line R-W. On this line, the retrograde condensation takes place and a part of the production fluid condenses. The production system performance is discussed in detail in Sub-section 5-2-2.

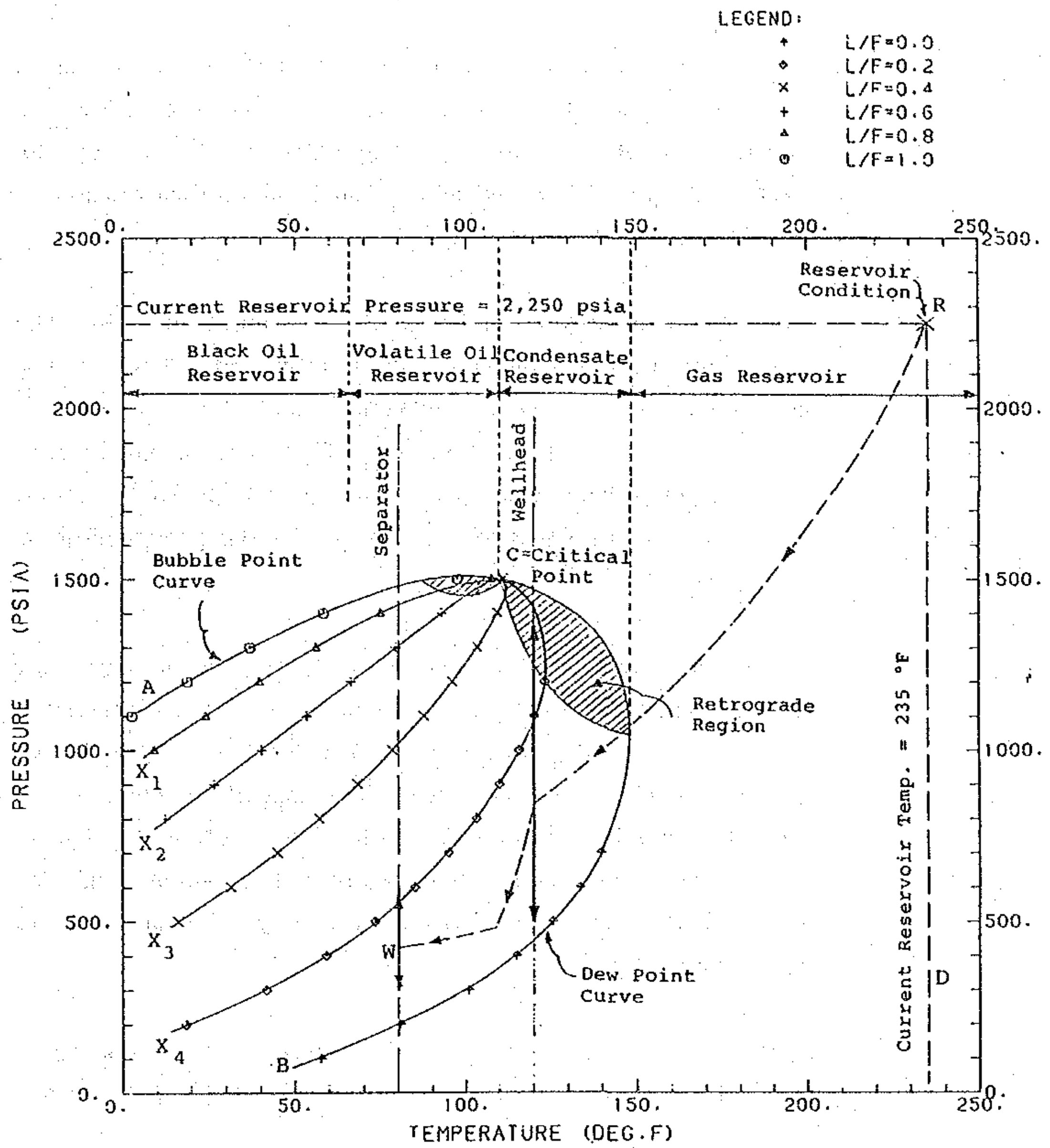


Figure 5-5 Pressure-Temperature Phase Diagram of Reservoir Fluid (for Assumed Gas Composition)

(c) Effect of Difference in Fluid Composition

As stated previously, the possibility can not be dismissed completely that the data of Table 5-6 does not show the composition of the reservoir fluid itself but that of the separated gas. The effect of such an interpretation is briefly discussed below.

The separated gas is in the state of equilibrium with the liquid (oil) in the separator and is saturated. That is, the separation temperature and pressure should be positioned on the dew point curve of the gas. Therefore, provided that the test separation temperature was 26.7°C (80°F), the pressure would have been about 14 kg/cm^{2}A (200 psia) according to Figure 5-5.

If the reservoir fluid is separated at the temperature and pressure, the separated gas composition (not the reservoir fluid composition) should agree with the composition listed in Table 5-6. The reservoir fluid composition which satisfies this condition cannot be deductively obtained. Therefore, a reservoir composition shown below which would be close to the actual one was obtained by trial-and-error as an example:

CO_2	C1	C2	C3	i-C4	n-C4	i-C5	n-C5	C6+
6.0	48.0	14.0	15.0	4.0	7.0	3.0	2.0	1.0

For reference, the pressure-temperature diagram for this composition is given in Figure 5-6. The amount of the heavier components in the composition is higher than that of the composition adopted in this study, so that the critical temperature is a little higher. However, the overall relationship with the reservoir temperature and pressure is not so different. The reservoir type in this case is also a single-phase gas reservoir. The effect on

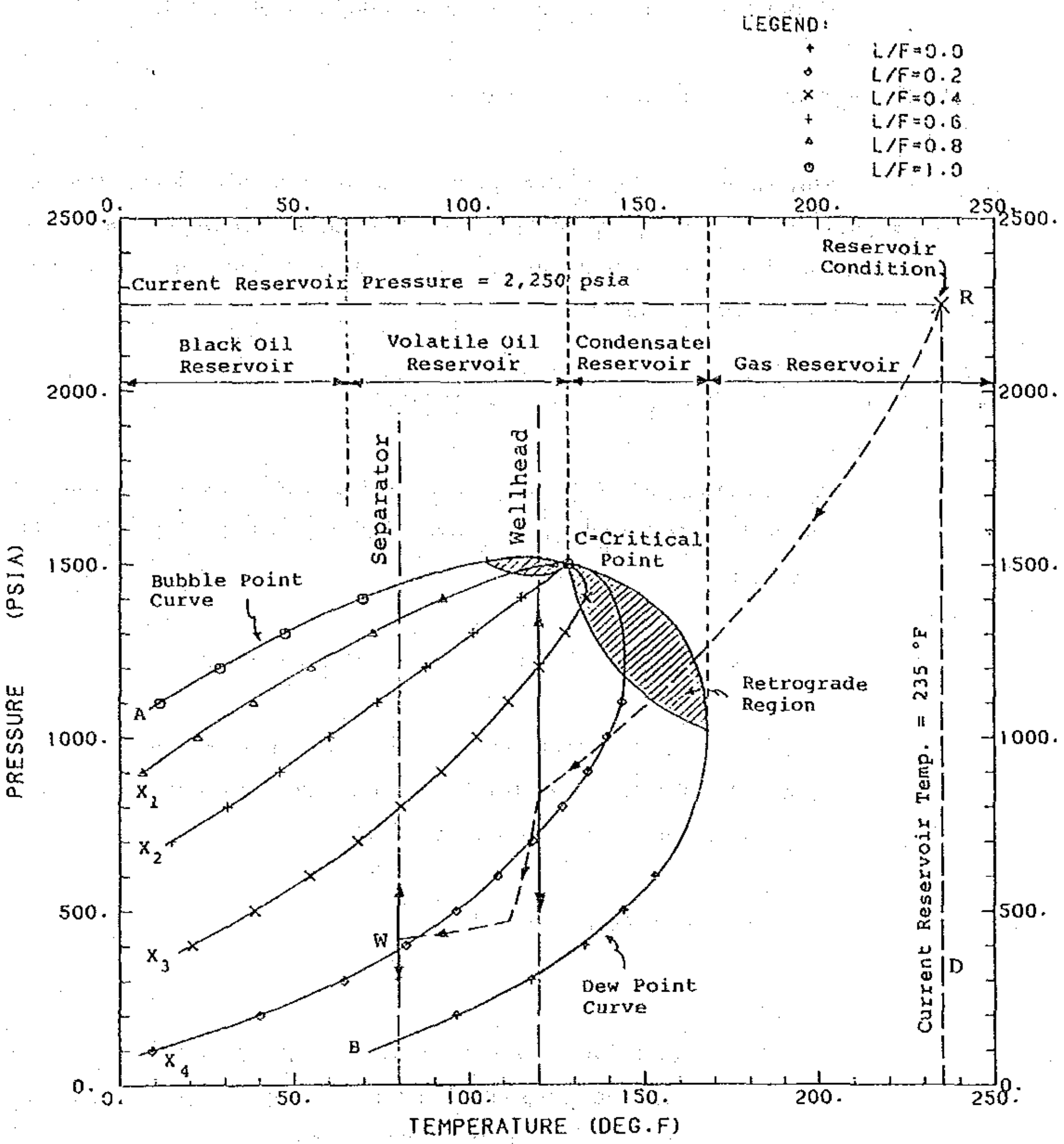


Figure 5-6 Pressure-Temperature Phase Diagram of Reservoir Fluid (for Reference)

the gas/condensate composition at the outlet of the gas oil separation plant is described in Sub-section 5-2-3.

(3) Production Performance of Reservoir

The objective reservoir was judged to be a single-phase gas reservoir. According to this judgment, production performance of the reservoir is estimated on the basis of reservoir material balance consideration.

(a) Material Balance in Gas Reservoir

Considering the reservoir as a vessel, the following equation regarding the material balance in the gas reservoir can be applied:

Gas produced = Initial gas - Remaining gas

Applying the equation of state to the above, the relationship between the cumulative gas production and the reservoir pressure is expressed as follows:

$$P/Z = (P_i/Z_i) \times (G_i - G_p)/G_i$$

where,

P : Reservoir pressure after producing G_p
Z : Gas compressibility factor after producing G_p
P_i : Initial reservoir pressure
Z_i : Initial gas compressibility factor
G_i : Original gas-in-place
G_p : Cumulative gas produced

The above equation can be applied under the following assumptions:

1) The gas reservoir has no water-drive mechanism.

2) There is no contribution of associated gas in the reservoir.

3) There occurs no retrograde condensation in the reservoir.

Because the original reservoir conditions are unknown, the current conditions shown below are regarded as the initial conditions.

P_i = 158 kg/cm^{2}A (2,250 psia)
G_i = Approx. 1.40 x 10^9 m^3 (49.7 BSCF)

Reservoir pressure decline can be estimated by calculating the gas compressibility factors which correspond to the declining reservoir pressures. The result is shown in Figure 5-7.

The recovery factors for volumetric gas reservoirs will range from 80 to 90% if there is no water-drive mechanism. If this is applicable, the recoverable reserve will be approximately 1.13 - 1.27 x 10^9 m^3 (40 - 45 BSCF) and the abandonment pressure will be approximately 21 - 42 kg/cm^{2}A (300 - 600 psia). However, these figures are approximate value, and further considerations will be given in the production systems analysis, described in Sub-section 5-2-3.

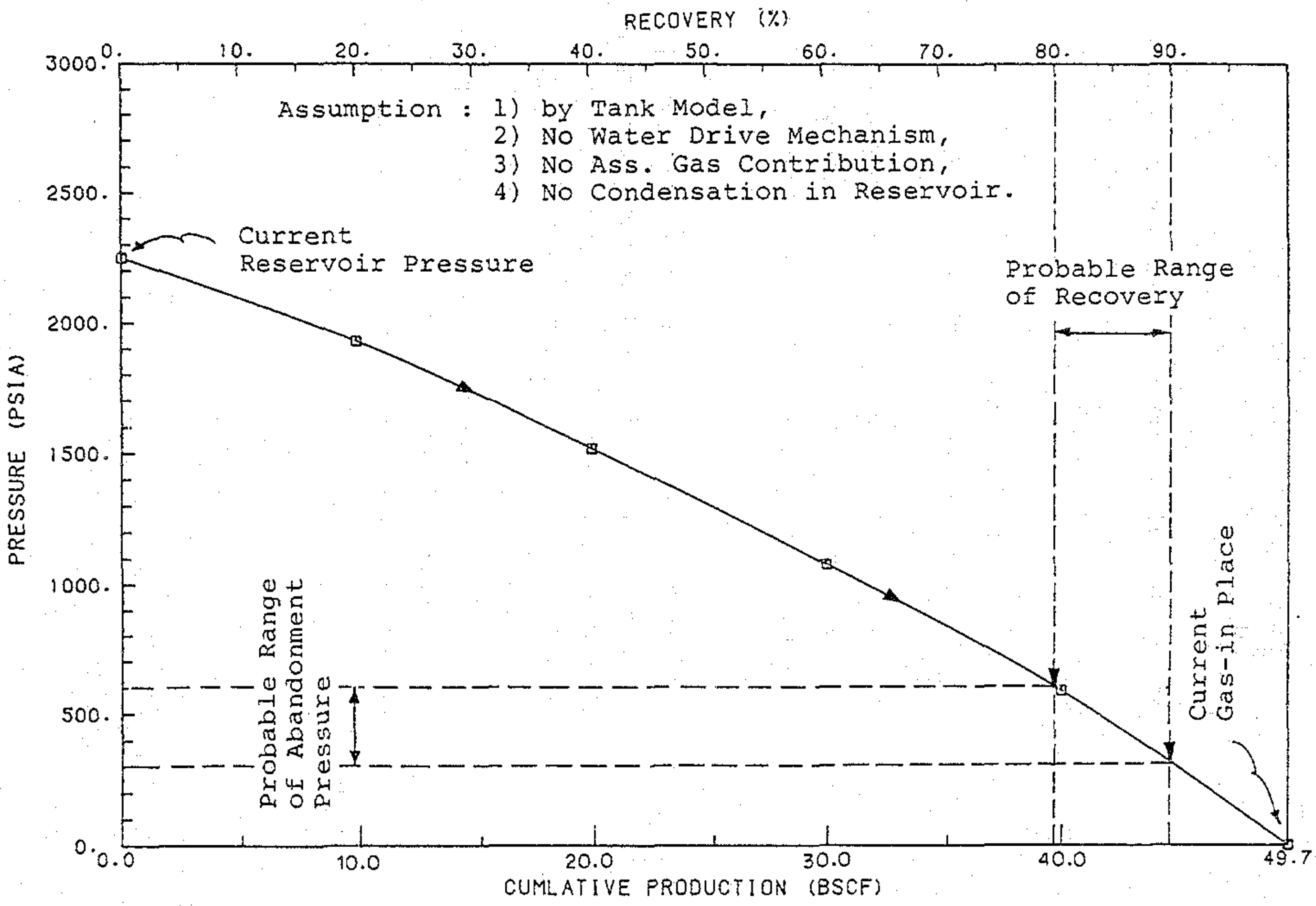


Figure 5-7 Reservoir Pressure Decline Curve

(b) Production Performance Prediction

In the above, reservoir pressure profile was correlated to the cumulative gas produced. The correlation between the profile and time, however, depends on the production plan. The profiles of the reservoir pressure decline along with time are shown in Figure 5-8 for several daily average production rates. The working days for the gas production are taken as 365 days per year for the calculation. The project life is assumed to be 20 years in this project. From this diagram the reservoir pressure, the cumulative production and the remaining reserves at the end of the 20 year production are estimated as follows:

Daily Average Production		Reservoir Pressure		Cumulative Production		Remaining Reserves	
$10^3 m^3/d$	(MMSCFD)	kg/cm^2A	(psia)	$10^9 m^3$	(BSCF)	$10^9 m^3$	(BSCF)
57	2	123	1,750	0.41	14.6	0.99	35.1
85	3	101	1,440	0.62	21.9	0.79	27.8
113	4	78	1,110	0.83	29.2	0.58	20.5
142	5	53	760	1.03	36.5	0.37	13.2
170	6	23	330	1.24	43.8	0.17	5.9

Judging from the recoverable pressure range of the objective gas reservoir (see Sub-section 5-2-2 for details), there is a possibility that the gas production becomes unsustainable because of the depletion of the reservoir before the end of the plant life, if the daily average production is set at 170 x 10^3 m^3/d (6 MMSCFD) or higher.

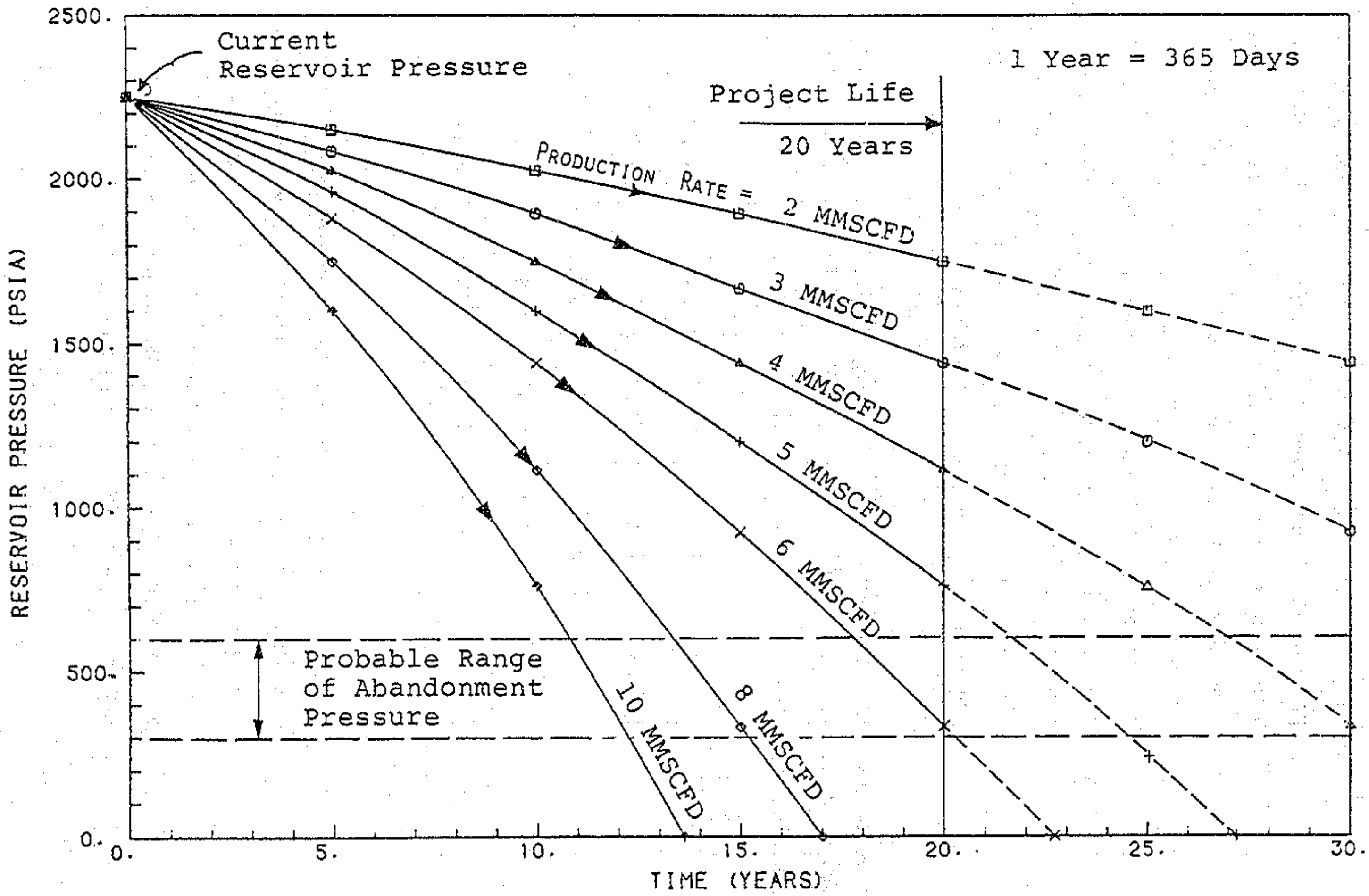


Figure 5-8 Reservoir Pressure Performance with Varying Production Rates

5-2-2 Production Systems Analysis

In this Sub-section, the well deliverability analysis is described as well as analysis of fluid flow in the wellbore and flowline, according to the gas reservoir performance described in Sub-section 5-2-1. In the analysis, the entire producing system, from the reservoir up to the separator, is treated as one system. The aim of this analysis is to clarify the following subjects by considering the system behavior throughout the entire gas production period:

1) Well deliverability and individual well production rate.
2) Presence of any bottle-neck in the production system.
3) Flowing temperatures and pressures in the production system.
4) Conditions which determine the range of the separator operating pressures.

(1) Production System Components

The reservoir gas flows into the wellbore and goes up to the wellhead through the tubing. Then, the gas is collected to the manifold through the individual flowline, and is sent to the high-pressure separator. The production rate is adjusted by applying proper back-pressure to the well with a choke provided at the wellhead assembly. To clarify the behavior of the entire production system, following must be evaluated:

1) Gas well deliverability (fluid flow to the bottom-hole through porous reservoir rock).
2) Fluid flow in tubing.
3) Pressure and temperature drops at choke.
4) Fluid flow in flowline.

In the following, characteristics of each system component and its principal data (including assumed/estimated values) are summarized:

(a) Reservoir

Type

: Gas reservoir

Datum depth

: 1,770 m (5,800 ft)

Average temperature

: 113°C (235°F)

Average pressure

: 158 kg/cm^{2}A (2,250 psia)

Fluid composition

: See Sub-section 5-2-1.

Gas well deliverability

: See Item (2) of this Sub-section.

(b) Well

The reservoir gas flows into the wellbore, and then goes up through the tubing in the state of gas single-phase in the deeper portion. There appears liquid condensate to turn into the vapor-liquid two-phase state while the gas goes further up through the tubing with decrease in both pressure and temperature (See Figure 5-5). As the method for estimating the behavior of the vertical two-phase flow in the well (relation between the flow rate and the pressure), various empirical correlations have been proposed. Among them, the Hagedorn & Brown's correlation, which is often used, was applied to the calculation, since there is no measured data which can be used for selecting any particular correlation. The fluid properties of the two-phase system and their changes according to pressure and temperature, which are necessary for the calculation, are obtained by using general correlations based on the fluid composition. Other principal data are as follows:

Well completion

: Single-layer tubing completion

Tubing diameter

: 2-7/8 inch (nominal size)

Well depth

: 1,770 m (5,800 ft)

(c) Wellhead Choke

In general, the production rate of naturally flowing well is adjusted by applying proper back-pressure with a choke. The reasons are as follows:

1) The production rate of individual well must be controlled in accordance with the most feasible reservoir production plan.

2) Excessive draw-down will cause the caving of reservoir rocks, sand entry or water coning in the vicinity of the bottom-hole. The draw-down means the difference between the static reservoir pressure and the flowing bottom-hole pressure. In other words, the draw-down is the pressure loss caused by the fluid flow through porous reservoir rock to the bottom-hole. Therefore, the larger the draw-down, the larger the production rate.

3) When a reservoir pressure is high and pressure difference across a choke is very high, gas velocity through the choke reaches to sonic velocity and gas production rate will become constant regardless pressure variation of the down stream side.

(d) Flowline

After its pressure reduced by the choke, the well fluid flows through the flowline in the state of two-phase. For the calculation of the behavior of the horizontal two-phase flow, Dukler & Eaton's correlations were used. The principal data on the calculation are given below. The average flowline length was obtained by excluding those for dry-holes (see Table 5-2).

Flowline diameter: 4 inch
Average flowline length: 1,200 m (3,880 ft)

(e) Separator

The operating pressure of the separator is taken as a parameter. The design pressure of the high-pressure test separator is approximately 39 kg/cm^2G. Taking account of the rating of the block valve at the flowline inlet and the required delivery pressure of the separated gas and condensate altogether, the proper range of the separator operating pressure is considered to be 20 - 40 kg/cm^2G.
The optimum pressure is to be decided by examining the entire system including the LPG recovery plant and the natural gas pre-treatment plant. In this study, cases with following separator operating pressures were set to determine the feasible range from the viewpoint of production system performance.

Separator operation pressure: 20/30/40/50/70 kg/cm^2G.

(2) Gas Well Deliverability

(a) Current Gas Well Deliverability

When the gas in the reservoir flows into the well, pressure loss, called pressure draw-down, takes place and the flowing bottom-hole pressure becomes lower than the static reservoir pressure. The following empirical equation based on Darcy's Law (a law regarding the fluid flow in porous media), which correlates the gas production rate, reservoir pressure and the flowing bottom-hole pressure, can be applied to steady state gas flow.

$$Q = C\,(P_R^{\,2} - P_{wf}^{\,2})^n$$

where

- Q : Gas production rate (at standard condition)
- P_R : Static reservoir pressure
  = 158 kg/cm²A (2,250 psia)
- P_{wf} : Flowing bottom-hole pressure
- C : Flow coefficient
- n : Exponent (normally, $n = 0.5 - 1.0$)

The gas well deliverability would be exactly determined by applying the above equation, if the value of flowing bottom-hole pressure and production rate were known at two or more points. However, only the following are given:

Absolute open flow potential
: 1.4×10^5 m³/d (5 MMSCFD)

Average deliverability range
: $5.7 - 8.5 \times 10^4$ m³/d (2 - 3 MMSCFD)

The absolute open flow potential, or AOF, is an index indicating the gas production rate that would occur if there could be no back pressure applied at the bottom-hole ($P_{wf} = 0$). The data can be expressed as follows:

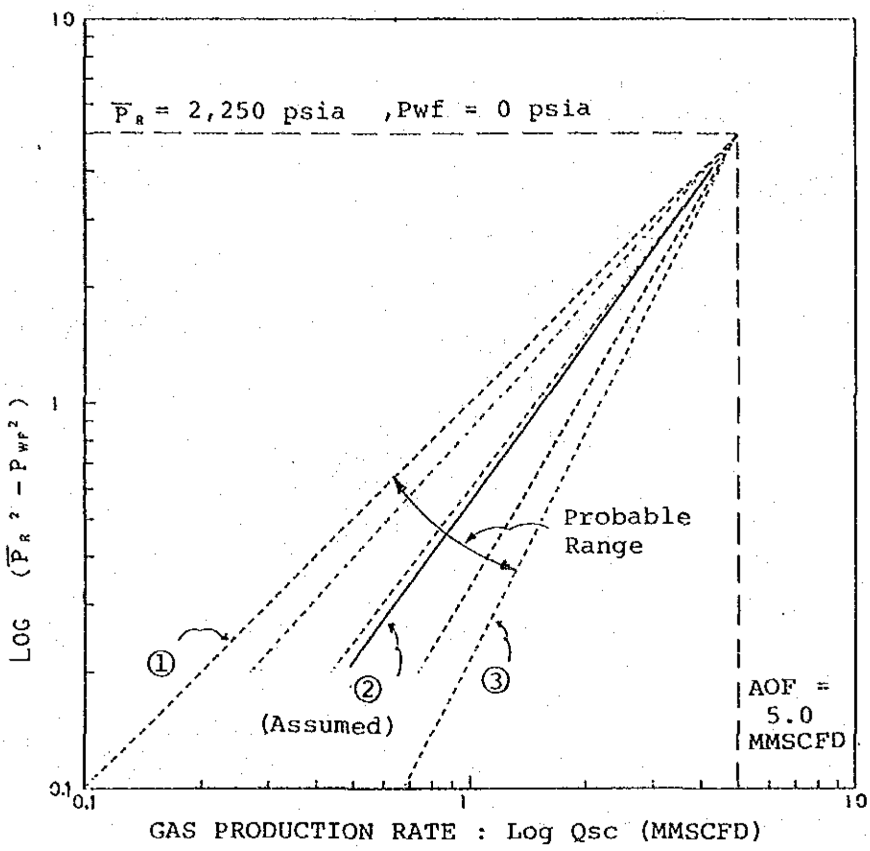
$$AOF = CP_R^{2n} = 1.4 \times 10^5 \ m^3/d \ (5 \ MMSCFD)$$

The exponent "n" in the gas well deliverability equation is the inverse of the straight line gradient when the gas flow rate Q and ($P_R^2 - P_{wf}^2$) are plotted on a logarithmic co-ordinates (see Figure 5-9). The exponent will fall normally between 1.0 (in the complete laminar flow state) and 0.5 (in the complete turbulent flow state). Therefore, the well deliverability curve of this gas reservoir could be located between the most outer broken line indicated in the figure.

As for the average deliverability range, it lacks the required information such as a flowing bottom-hole pressure which corresponds to a certain production rate, or conditions to estimate the pressure (such as the wellhead pressure, the surface separator pressure, etc.).

As is mentioned in the above an excessive draw-down causes damage to the well, so that the draw-down is usually held at approximately 10 - 20% of the static reservoir pressure. If this is true, the average deliverability range mentioned above will correspond to the flowing bottom-hole pressures of approximately 127 - 140 kg/cm^{2}A (1,800 - 2,000 psia). Figure 5-9 shows three possible variations of the gas well deliverability for this range by broken lines.

Figure 5-10 shows the well deliverability as the relation between the production rate and the flowing bottom-hole pressure (which is called IPR curve, or Inflow Performance Relationship Curve). The relative deviation of the curves shown in Figure 5-10 is not considered so significant, comparing with the accuracy of the other basic data.

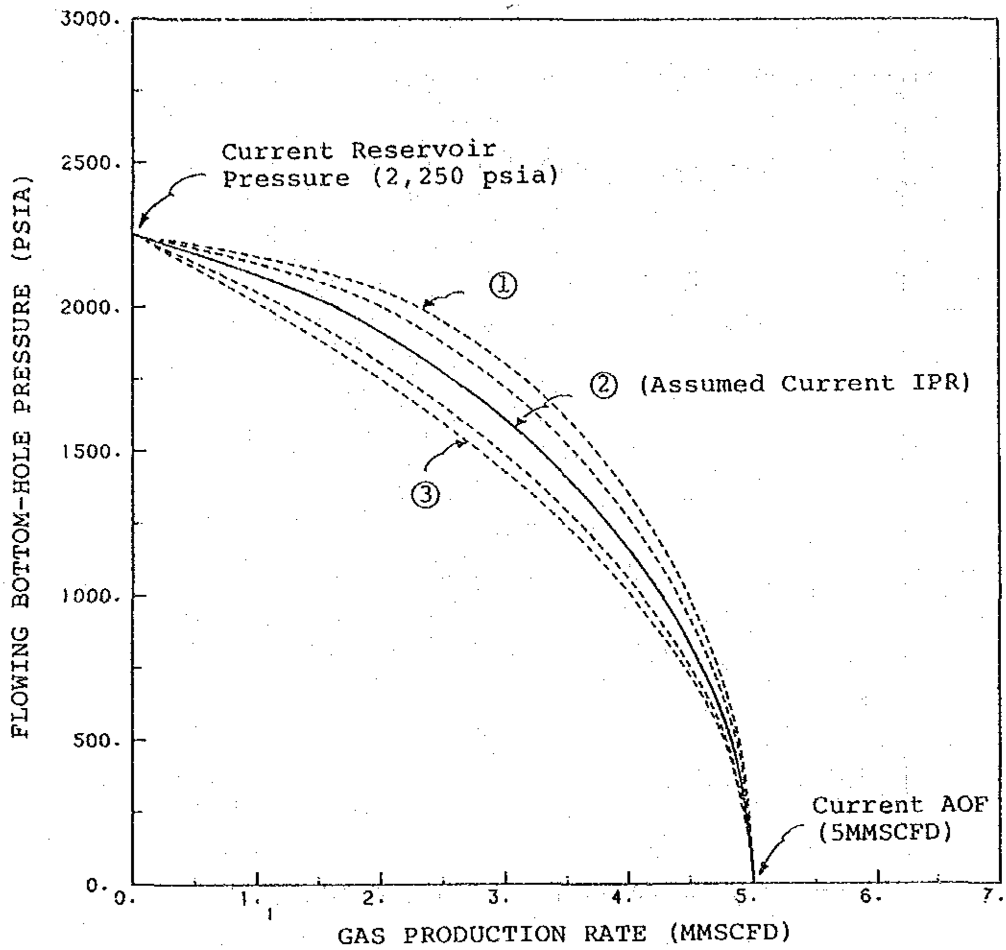


Gas Equation: $Q = C(\overline{P}_R{}^2 - P_{wf}{}^2)^n$

where, $\overline{P}_R$ = Reservoir Pressure (psia)
P_{wf} = Flowing Bottom-hole Pressure (psia)

Line No.	C	n	Draw-down (at 2 MMSCFD)
①	2,222	0.5	approx. 8 %
②	80.7	0.715	approx. 15 %
③	0.9877	1.0	approx. 25 %

Figure 5-9 Gas Deliverability Curves



Gas Equation: $Q = C(\overline{P}_R{}^2 - P_{wf}{}^2)^n$

where, $\overline{P}_R$ = Reservoir Pressure (psia)

P_{wf} = Flowing Bottom-hole Pressure (psia)

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①	2,222	0.5	approx. 8 %
②	80.7	0.715	approx. 15 %
③	0.9877	1.0	approx. 25 %

Figure 5-10 Gas Inflow Performance Relationship (IPR Curve)

In conclusion, 15% draw-down is assumed for the well production rate of 5.7 x 10^4 m^3/d (2 MMSCFD). This gives the following equation for the current gas well deliverability:

$$Q\ (m^3/d) = 101.8\ \{158^2 - P_{wf}\ (kg/cm^2A)^2\}^{0.715}$$

In Figures 5-9 and 5-10, the relationship given by the above equation is shown by a solid line.

(b) Future Inflow Performances

Since the reservoir pressure declines as the production proceeds, the gas well deliverability also decreases. In the production systems analysis, the entire period of the future gas production should be considered. On the basis of the current IPR assumed in the above, the future IPR curves were estimated as shown in Figure 5-11. For this estimation, Darcy's Law was applied. In the actual production, there are possibilities of well troubles due to formation of water coning, sand entry, etc. Therefore the future deliverability tends to become lower than the estimated.

The principal factor which controls the change of the future IPR curves is the decline of the reservoir pressure. As stated in Sub-section 5-2-1, the reservoir pressure shows different decline trends according to different production rates (see Figure 5-8). Therefore, the change of the IPR curve also depends on the field-total production rate, or the production plan, as shown in Figure 5-11. In the following discussion, the field production rate is taken as a parameter, as shown in the figure.

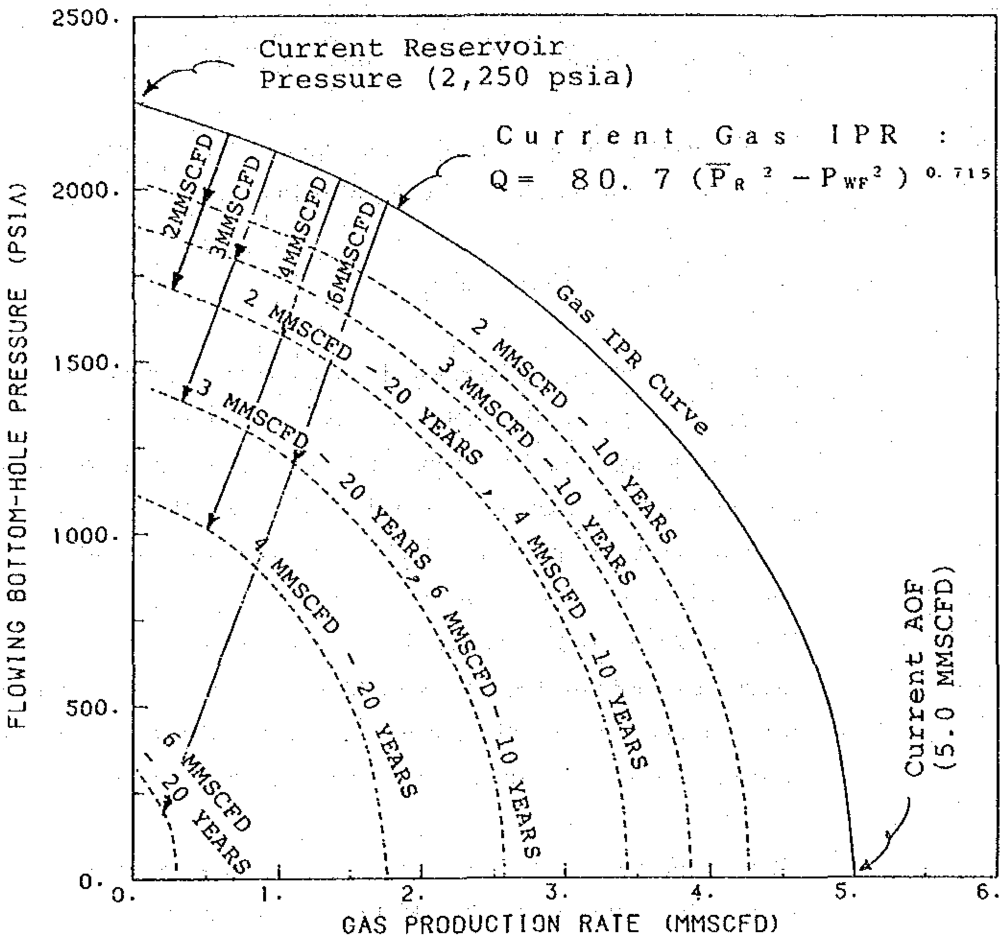


Figure 5-11 Future Gas IPR Curve with Varying Production Rate

(3) Production System Deliverability

(a) Procedure for Analysis

The deliverability (production potential) of the entire production system means the maximum possible production capacity of a naturally flowing well when not restricted by a surface choke. The factors which determine the production rate are the separator operating pressure and the flowing pressure losses in the tubing and flowline. Figure 5-12 shows a schematic of the system analysis. In this analysis, the node, the solution point of the pressure analysis, is placed at the bottom of the well. The entire system is divided into two components and the performance in each component is evaluated; the inflow performance for the well to the node, and the pressure performance through the pipings from the separator to the node.
The solution procedure is summarized below:

1) Assume several flow rates and construct the IPR curve, which represents the flowing bottom-hole pressure during the production.

(Node Inflow Pressure)
= (Static Reservoir Pressure) - (Draw-down)

2) Assume the separator pressure and obtain the required flowing bottom-hole pressure to move the fluid through the vertical tubing and horizontal flowline to the separator. This curve, which represents the correlation between the production rates and the required bottom-hole pressures, is named TIP curve or Tubing Intake Pressure Curve.

(Node Outflow Pressure) = (Separator Pressure)
+ (Pressure Losses in Flowline)
+ (Pressure Losses in Tubing)

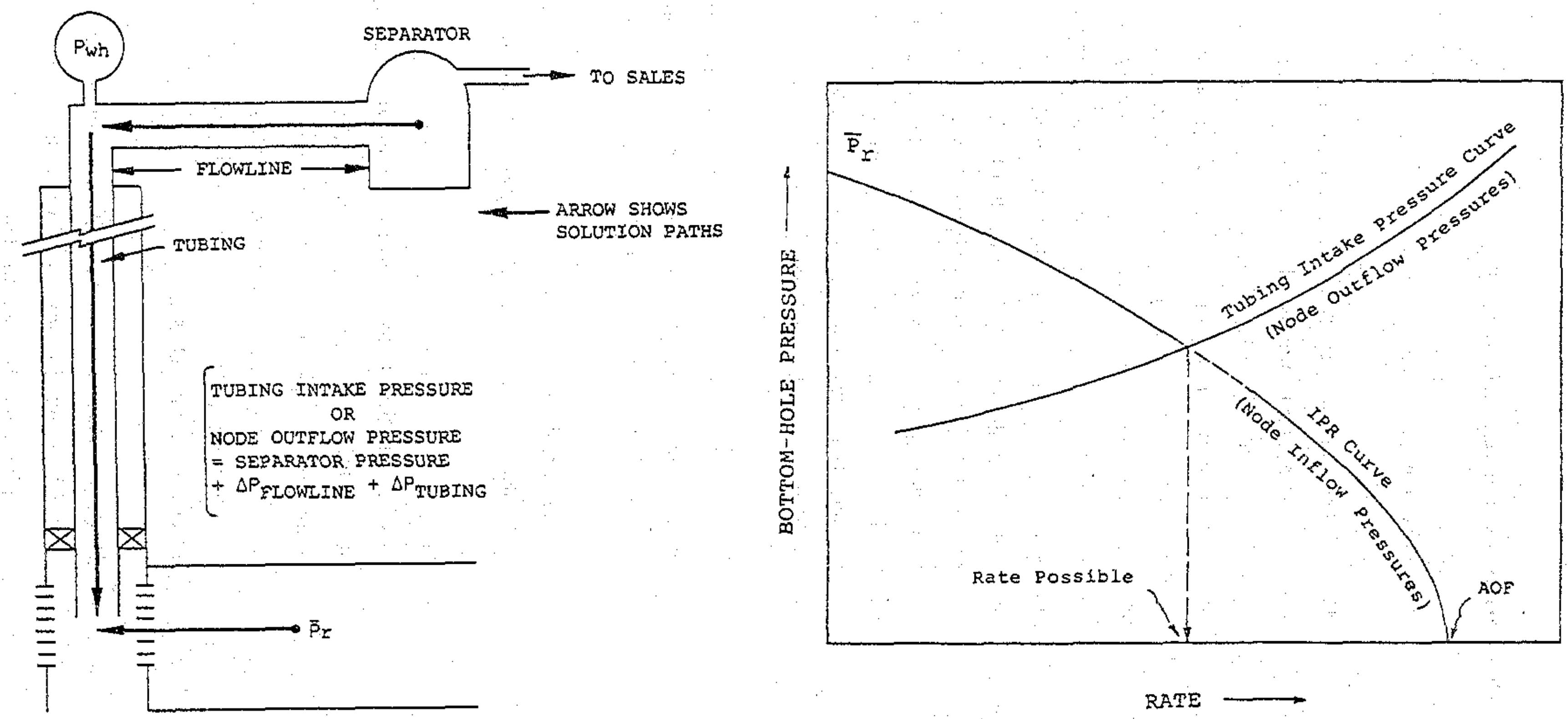


Figure 5-12 Schematic Diagrams for System Deliverability Analysis (Bottom-hole Node Solution)

3) Plot the IPR curve vs TIP curve on a graph. The intersection of these two curves shows the maximum production rate possible for the system.

(b) Consideration on the Results

The deliverability of the existing production system was analyzed according to the above procedure. The result of the analysis is shown in Figure 5-13. In the figure, five TIP curves are plotted for different separator operating pressures. The IPR curves are plotted for several kinds of production plan given in Figure 5-11 (limited to the IPR curves at current, 10th and 20th year after the start of production). In the following, considerations based on Figure 5-13 are described:

(i) Separator Operating Pressure

Decline rate of the reservoir pressure depends on the production plan of the whole field, as is given by the IPR curves in the figure. Provided that the production rate is set at 85 x 10^3/d (3 MMSCFD) or less, individual wells have sufficient system deliverability throughout the production term (20 years) with the separator pressure range of 20 to 50 kg/cm^{2}G. If the field rate is set at 113 x 10^3 m^3/d (4 MMSCFD), the separator pressure would have to be set at 40 kg/cm^{2}G or less, taking account of the final years of the production term.

The required gas production rate for this project will fall within the range of approximately 57 - 113 x 10^3 m^3/d (2 - 4 MMSCFD). Taking the entire production term into account, the proper separator pressure will fall within the range of 20 - 40 kg/cm^{2}G. Considering that many of the basic data

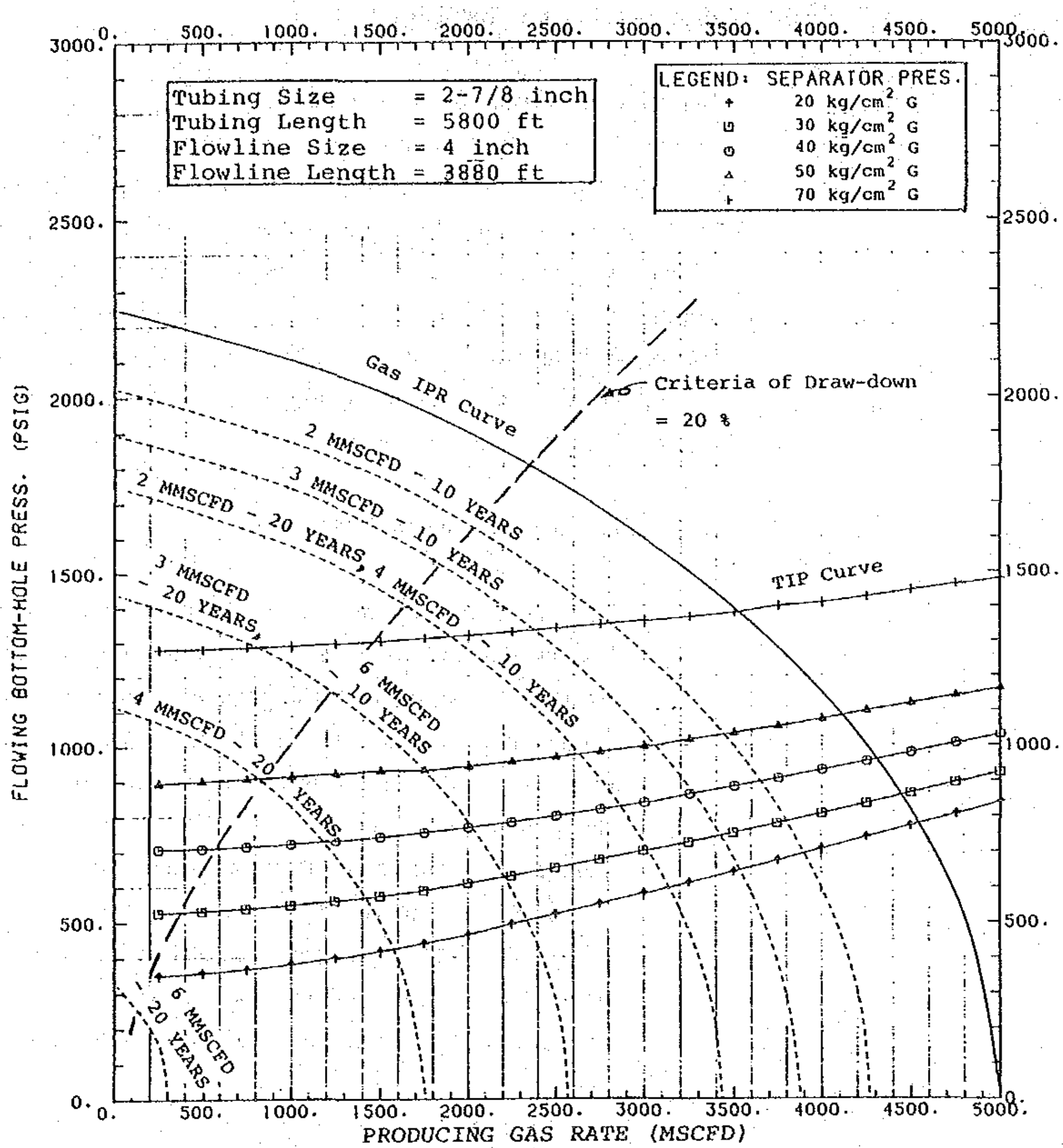


Figure 5-13 IRR vs. TIP Curves

used for the calculation are assumed and not technically verified, it would be safer to set the separator pressure, as the plant design parameter, as low as possible within the range.

(ii) Field-Total Production Rate

When the field total production rate is set at approximately 57 - 113 x 10^3 m^3/d (2 - 4 MMSCFD), pressure of gas reservoir declines slowly and it will be possible to maintain the estimated production rate throughout the production term (20 years).

However, if the production plan exceeds 170 x 10^3 m^3/d (6 MMSCFD), the well deliverability will drop to about a half of the initial deliverability within 10 years, and 20 years after, the wells will not be able to sustain natural flow even by lowering the separator pressure.

(iii) Well Production Rate

The well deliverability discussed here is the production rate which would occur if the well is produced without choke. The draw-down would be excessive in such a case. Therefore, it is necessary to apply a proper back-pressure by means of a choke in the actual operation. The production criteria for each well is derived from the optimum reservoir production plan. Here, however, 20% draw-down is assumed from the viewpoint of sound production practice. The proper production rates per well under this production criteria and corresponding values of the flowing bottom-hole pressures are shown in Figure 5-13. According to the figure, the individual well production rate at initial stage is a little more than approximately 68 x 10^3 m^3/d (2.4 MMSCFD). It declines corresponding to the reservoir pressure decline. For the reservoir pressure of 70 kg/cm^2A (1,000 psia), the well production rate is a little more than approximately 20 x 10^3 m^3/d (0.7 MMSCFD). This well behavior is discussed later in detail together with the estimation of the wellhead pressure.

(4) Production System Performance

(a) Procedure for Analysis

As stated in the above, the well production rate is controlled by the wellhead choke in the actual production operation. The system performance in the actual production is analyzed under the following procedure by taking the choke as the node for pressure analysis (see Figure 5-14):

1) The pressure drop from the reservoir to the choke inlet is calculated and the relation between the production rate and the flowing wellhead pressure, which is called V-IPR curve or Vertical IPR Curve, is obtained.

(Node Inflow Pressure) = (Static Reservoir Pressure)
 - (Draw-down at Bottom-hole)
 - (Pressure Losses in Tubing)

2) Assuming the separator pressure and proper flow rates, the relation between the separator pressure and the choke outlet pressure required for the fluid to reach the separator through the flowline, which is called CP curve, or Choke Performance Curve, is obtained.

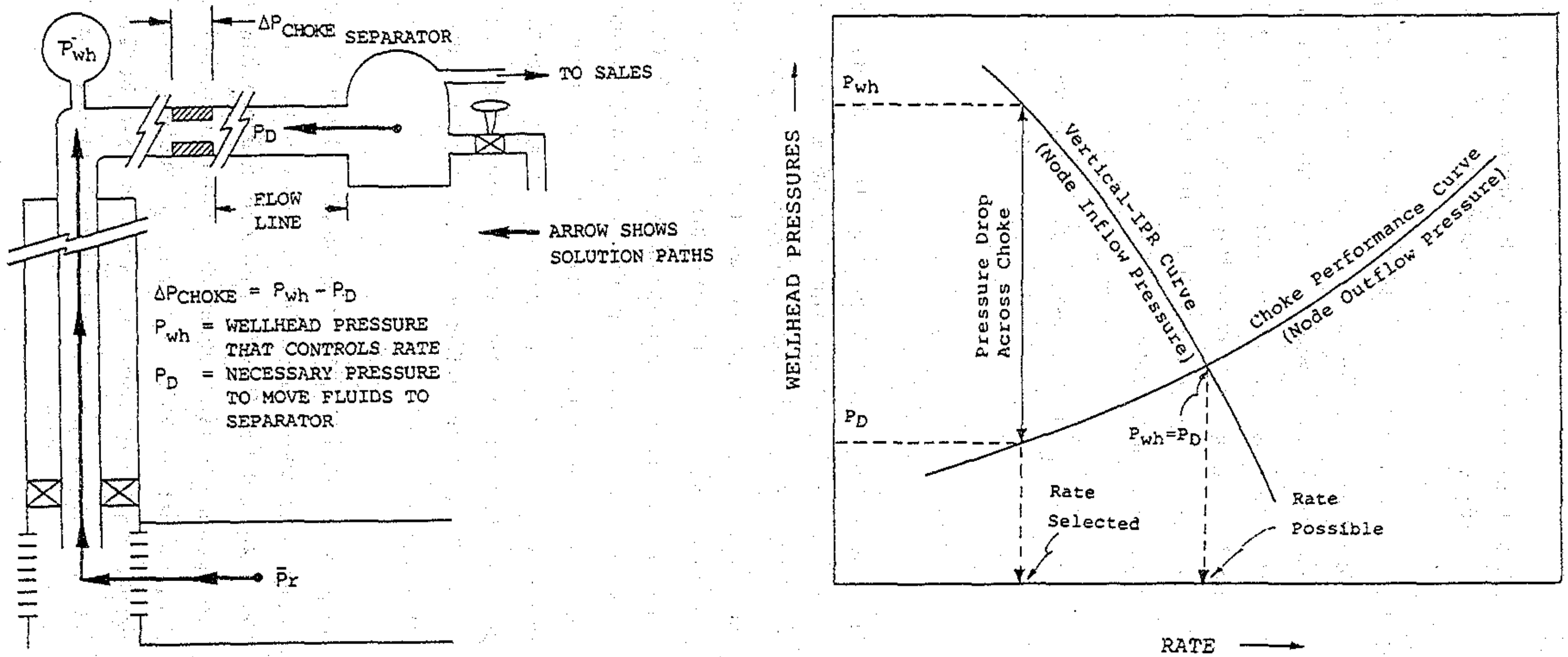


Figure 5-14 Schematic Diagrams for Production System Performance Analysis (Wellhead Node Solution)

(Node Outflow Pressure) = (Separator Pressure)
+ (Pressure Losses in Flowline)

3) Plot the V-IPR curve vs CP curve on a graph. The intersection of these two curves show the maximum production rate possible for the system when the production is conducted without using a choke. The pressure drop caused by the choke is obtained as the pressure difference given by the two curves for the same production rate.

(Pressure Drop at Choke) = (V-IPR Curve)
- (CP Curve)

(b) Consideration on the Results

The production performance of the existing production system was analyzed according to the above procedure, and the result of the analysis is shown in Figure 5-15. In this figure, the V-IPR curve moves downward as time goes by (*the V-IPR curves at current, 10th and 20th year after the* start of production), corresponding to the same production plans as those assumed in the deliverability analysis. On the other hand, there are five CP curves drawn by taking the separator operating pressure as the parameter. The pressure loss in the flowline is quite small and there is little effect of the difference in production rate in all the CP curves.

Same as in Figure 5-13, the upward sloping broken line in Figure 5-15 shows the relation between the proper production rate and the flowing wellhead pressure (the production rate guiding curve) under the assumption that the drawdown is set at 20% as a well production criterion. The proper production rate at any time can be given by the intersection of the production rate guiding curve and the

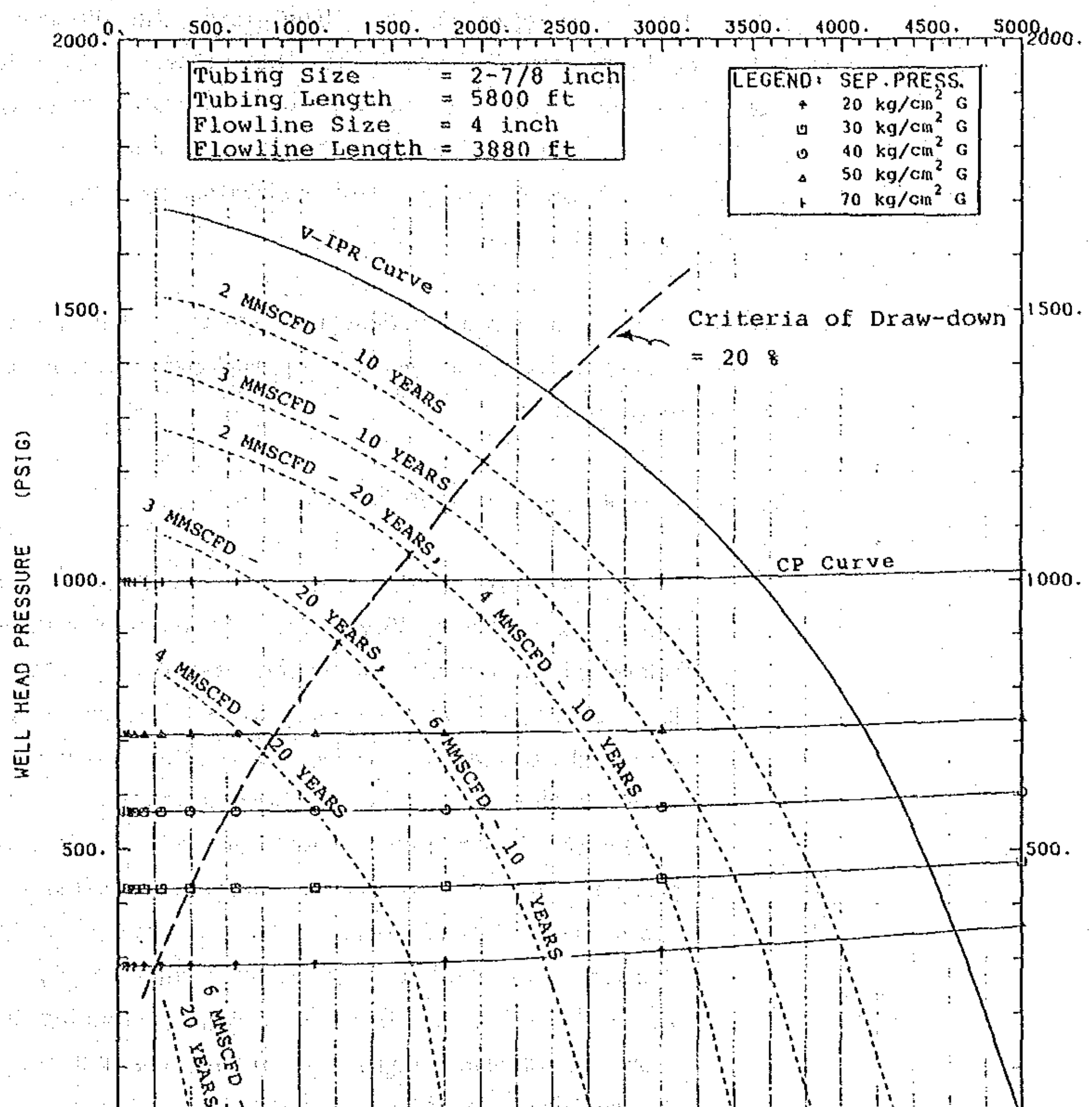


Figure 5-15 V-IPR vs. DP Curves (Tubing Size = 2-7/8 inch)

V-IPR curve of particular production plan and time. The pressure given by the V-IPR curve at any production rate is the flowing wellhead pressure at the choke inlet at that production rate, while that given by the CP curve is the pressure at the choke outlet. Therefore, the difference between these two pressures is the pressure reduction caused by the choke. If the CP curve is higher than the V-IPR curve, that production rate is exceeding the well production capacity against the separator pressure indicated by the CP curve.

Provided that the actual production operation follows the above production criteria, possible operational problems are evaluated as follows:

(i) *Flowing Condition and Tubing Size*

In a gas well, if the flowing gas velocity in the tubing is too low, liquid droplets (condensate and water) in the gas stream, which have precipitated in the tubing due to decrease in both pressure and temperature, can not move up to the surface and they accumulate in the wellbore gradually. If the production is continued under such a condition, the pressure loss in the tubing increases gradually to lower the wellhead pressure. As a result, the production rate decreases gradually, and finally the well stop flowing. In such a case, unloading (the work of clearing away the liquid accumulated in the wellbore) should be periodically conducted.

From this viewpoint, the existing 2-7/8-inch tubing might be a little too large for the planned gas production. All the existing facilities are thought to be designed for the purpose of producing crude oil and condensate, while the aim of this project is gas production. Therefore, there may be a possibility that the existing tubing size is too large. The following examinations are conducted to review this point.

The minimum flow velocity required for the liquid droplet (condensate or water) in the tubing to move up to the wellhead is calculated by a simple empirical equation. The comparison between the minimum velocity for continuous liquid removal and the gas flowing velocity in 2-7/8-inch tubing, at both the bottom-hole and wellhead, is shown in Figure 5-16. As per this figure, the liquid load-up would take place when the well production rate is 57×10^3 m^3/d (2 MMSCFD) or less, even if the fluid does not contain water droplet.

In addition, the result of the flow velocity examination for 1-inch tubing is shown in Figure 5-17 for comparison. In this case, production rate of approximately 23×10^3 m^3/d (0.8 MMSCFD) or more causes no liquid load-up even if a little amount of water is produced with the gas. However, with the smaller tubing, the pressure loss will be larger and the well deliverability will be reduced as a result. To see the influence, the well deliverability with 1-inch tubing was analyzed as shown in Figure 5-18 for a reference purpose.
According to the result, the well deliverability can be maintained at approximately 28×10^3 m^3/d (1 MMSCFD) by setting the separator pressure in the range of 20 to 40 kg/cm^2G, when the total field production rate is 85×10^3 m^3/d (3 MMSCFD) or less.

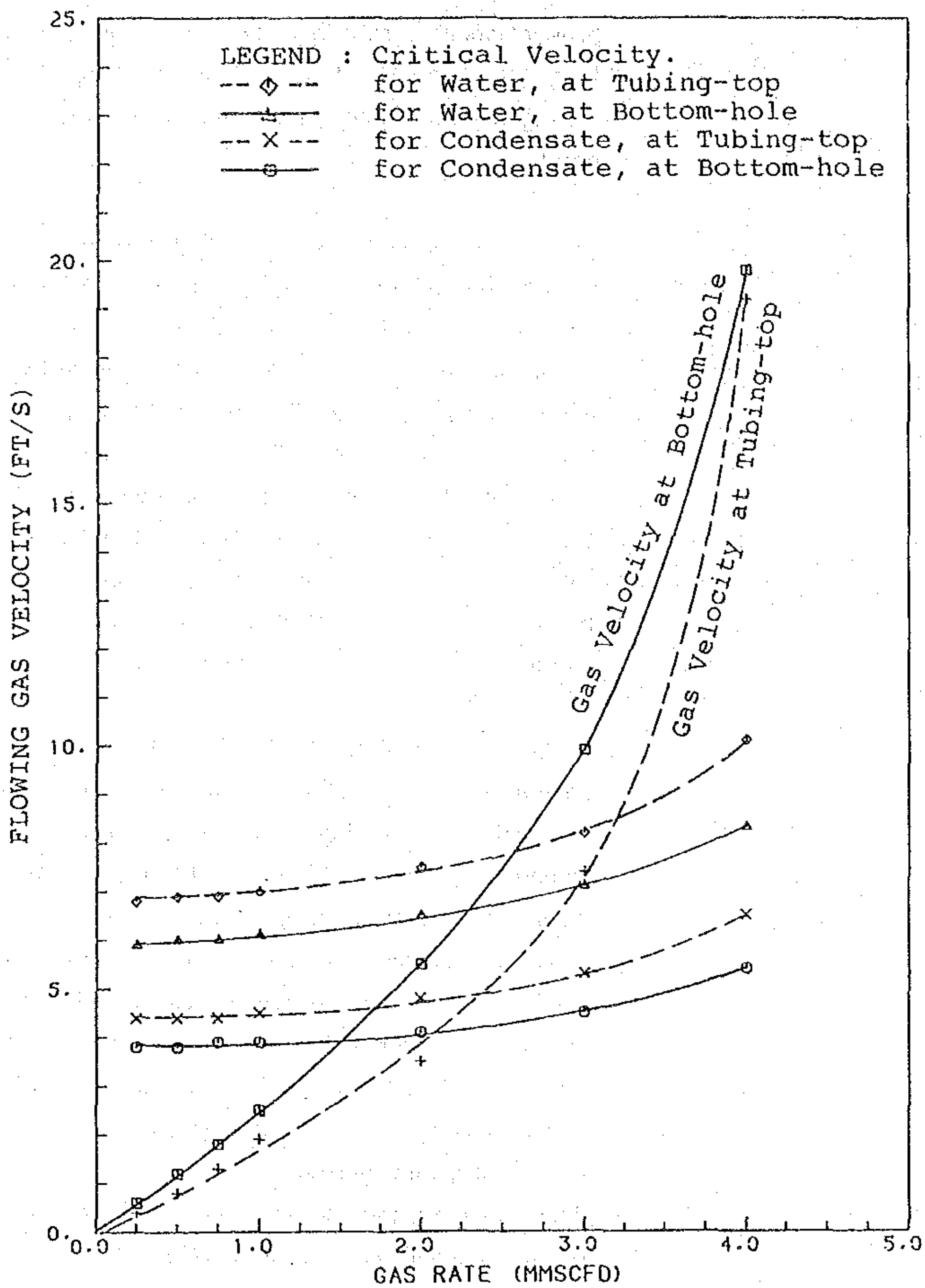


Figure 5-16 Gas Velocity through Tubing
(Tubing Size = 2-7/8 inch)

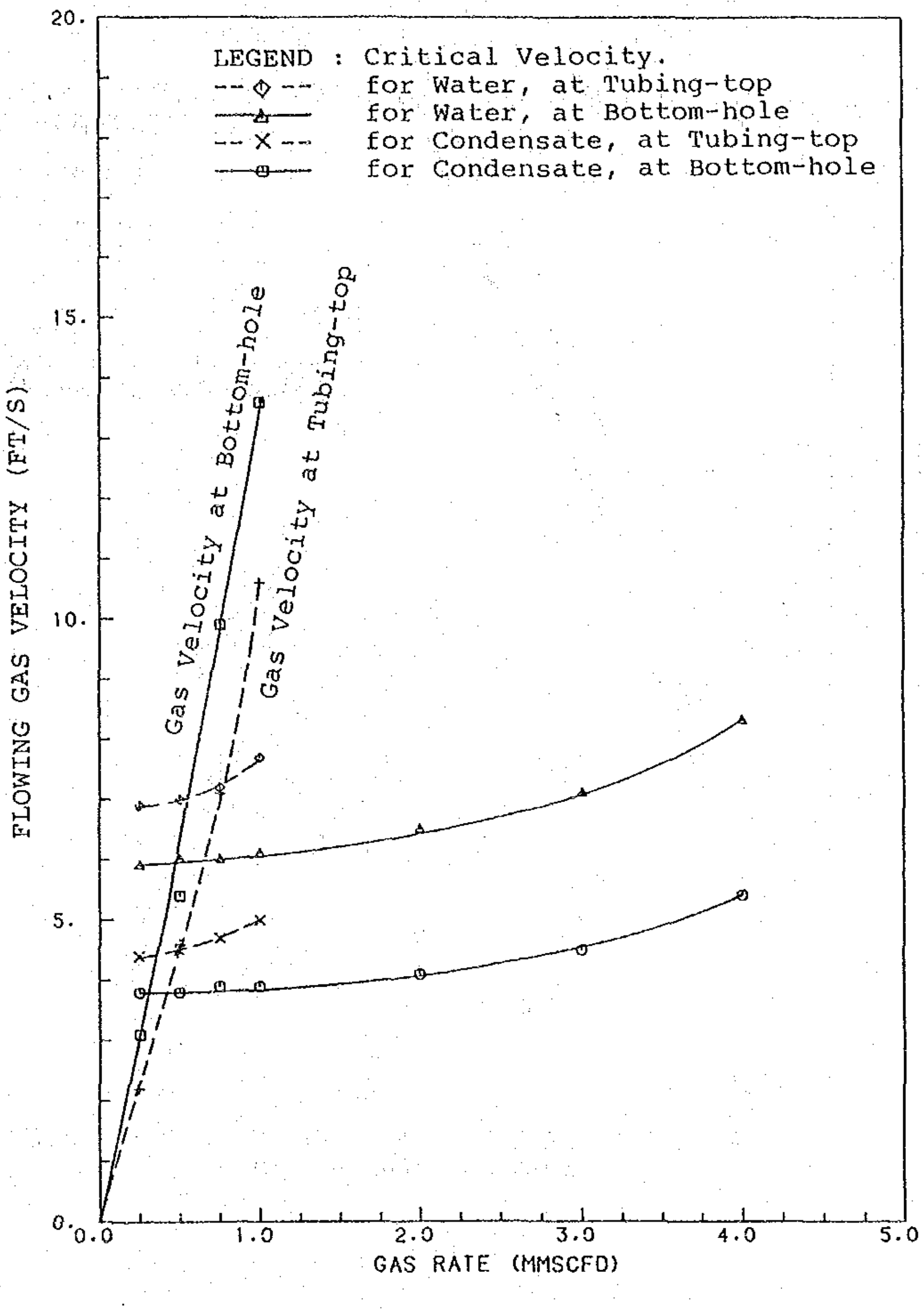


Figure 5-17 Gas Velocity through Tubing
(Tubing Size = 1 inch)

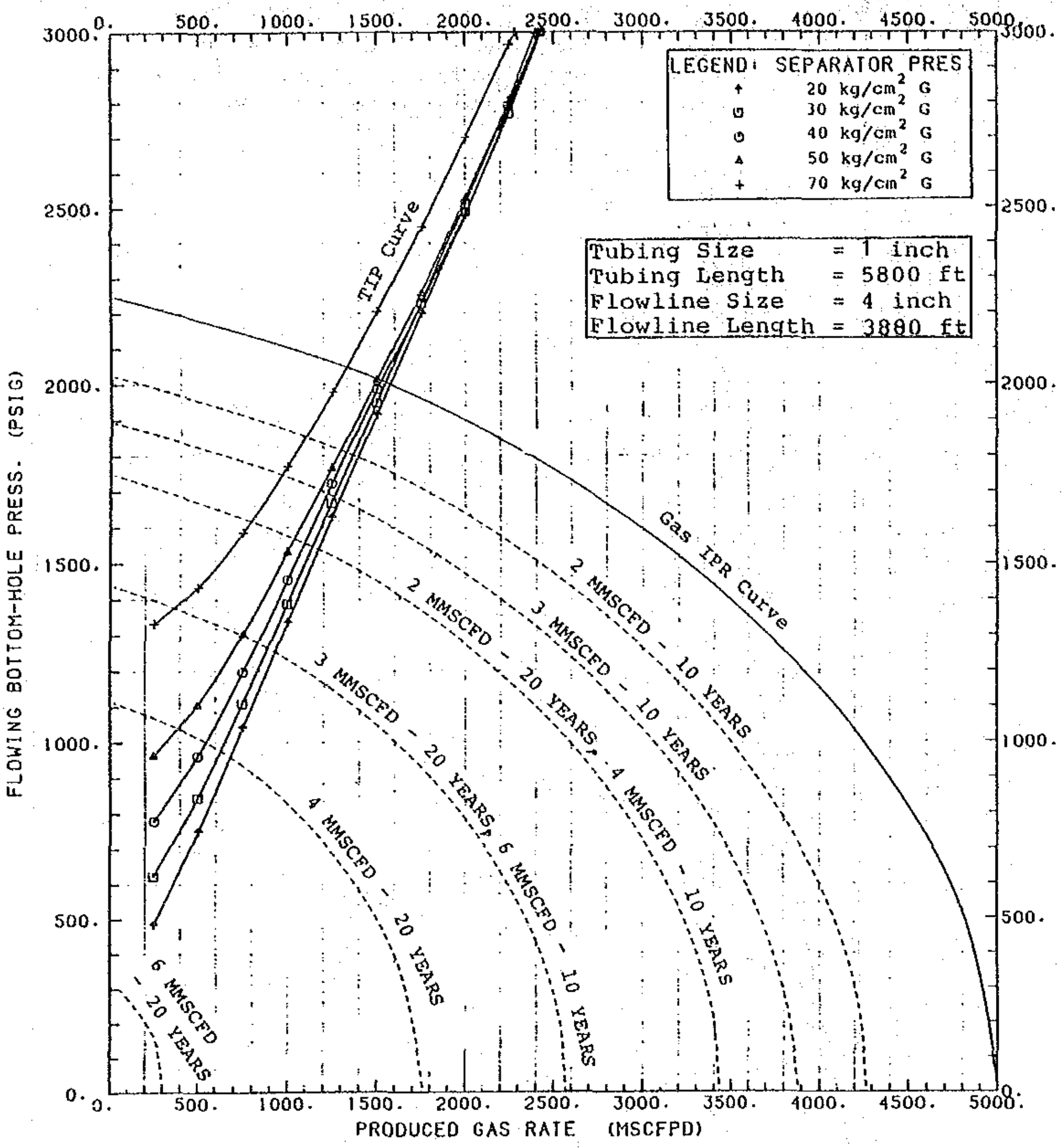


Figure 5-18 IPR vs. TIP Curves
(Tubing Size = 1 inch)

In conclusion, it can be said that the existing tubing size (2-7/8-inch) may be too large for the planned gas production, and there are some possibility of liquid build up in the tubing. Therefore, after close examination of the past production data, and if such phenomenon had been taken place in actual operation, it would be necessary to consider some kind of countermeasures to prevent such a phenomenon.

(ii) Gas Hydrate Formation at Choke

The pressure drop at the choke reaches its maximum at the initial stage of production when the wellhead pressure is the highest. If the pressure is sharply reduced by a choke, the fluid temperature drops due to the Joule-Thomson effect. The relation between the pressure drop and the temperature drop for the separator pressures of 20, 30 and 40 kg/cm^2G is shown in Figure 5-19. The maximum pressure and temperature drops for the above separator pressures, assuming the production criteria of 20% draw-down, are estimated as follow:

Separator Pressure	Pressure Drop		Temperature Drop	
kg/cm^2G	(kg/cm^2)	(psi)	(°C)	(°F)
20	70	1,000	37	66
30	64	910	29	53
40	55	780	22	40

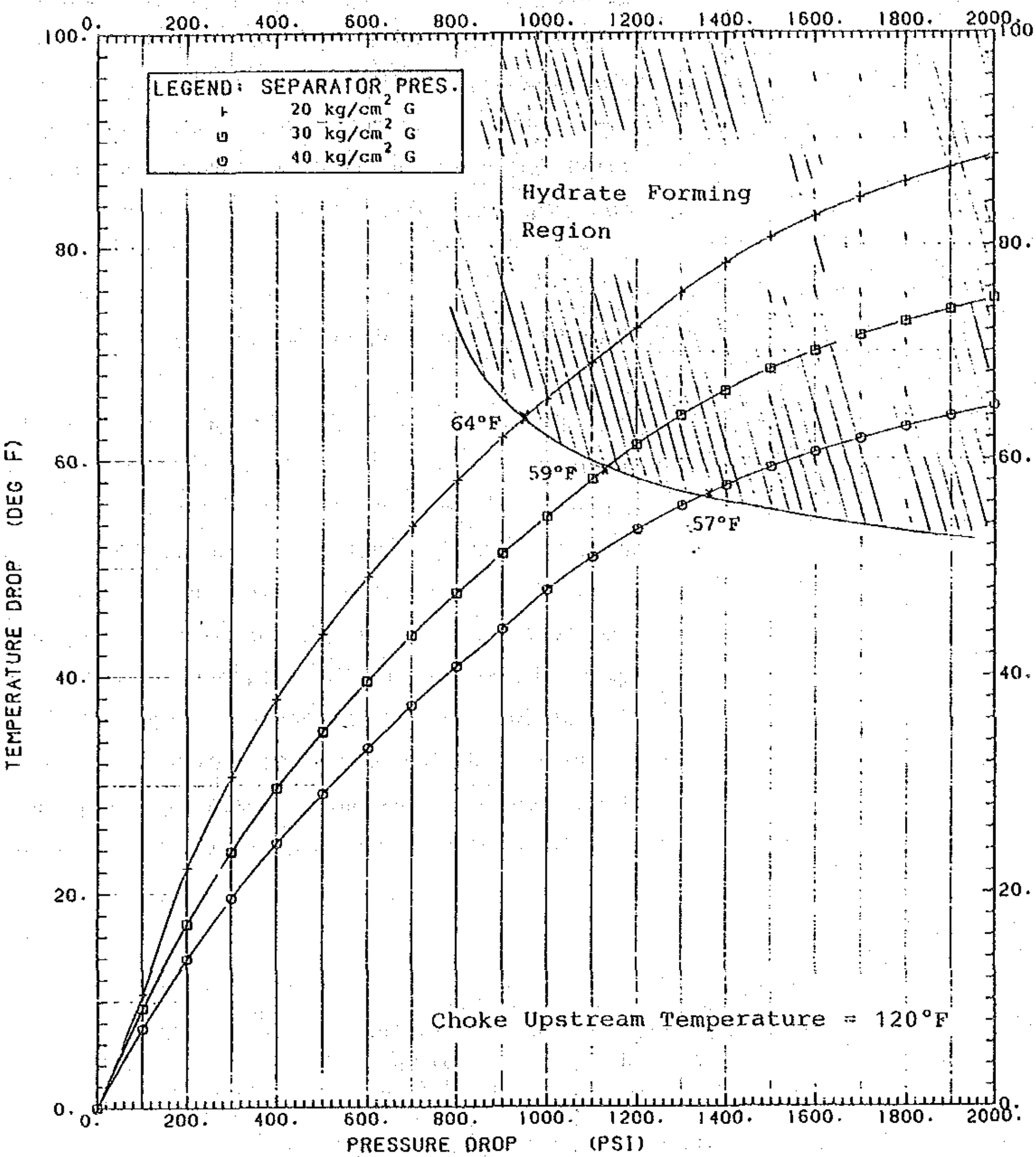


Figure 5-19 Pressure and Temperature Drops at Choke

Natural gases form a solid gas hydrate when they co-exist with liquid water under high pressures even if the temperature is 0°C or higher. Once the hydrate forms, it grows in tubing or flowline and may cause excessive flowing pressure loss, clogging within the pipe, etc. Gas producing wells commonly have a possibility of the hydrate formation, especially with a sharp temperature drop at the choke.

Figure 5-19 shows the hydrate forming region for a probable tubing-top temperature of 48°C (120°F).

Judging from this figure, during initial operating period when the reservoir pressures are high, there are possibility of hydrate formation and detail examination will be required.

(5) Performance Prediction for Individual Well Production

(a) Production Rate and Pressure Profiles

The predicted well production performance under the production criteria (draw-down 20%) is summarized in Figure 5-20, according to the analysis discussed in Item (4). The figure shows profiles of the well production rate, flowing bottom-hole pressure, flowing wellhead pressure at both the choke inlet and outlet, correlated with the reservoir pressure decline. These profiles can be correlated with time by assuming a production plan for the entire field.

The well production rates in the 10th and 20th year in accordance with the field production rates can be predicted as follows (the appropriate maximum production rate per well during initial production stage is approximately 68 x 10^3 m^3/d (2.4 MMSCFD):

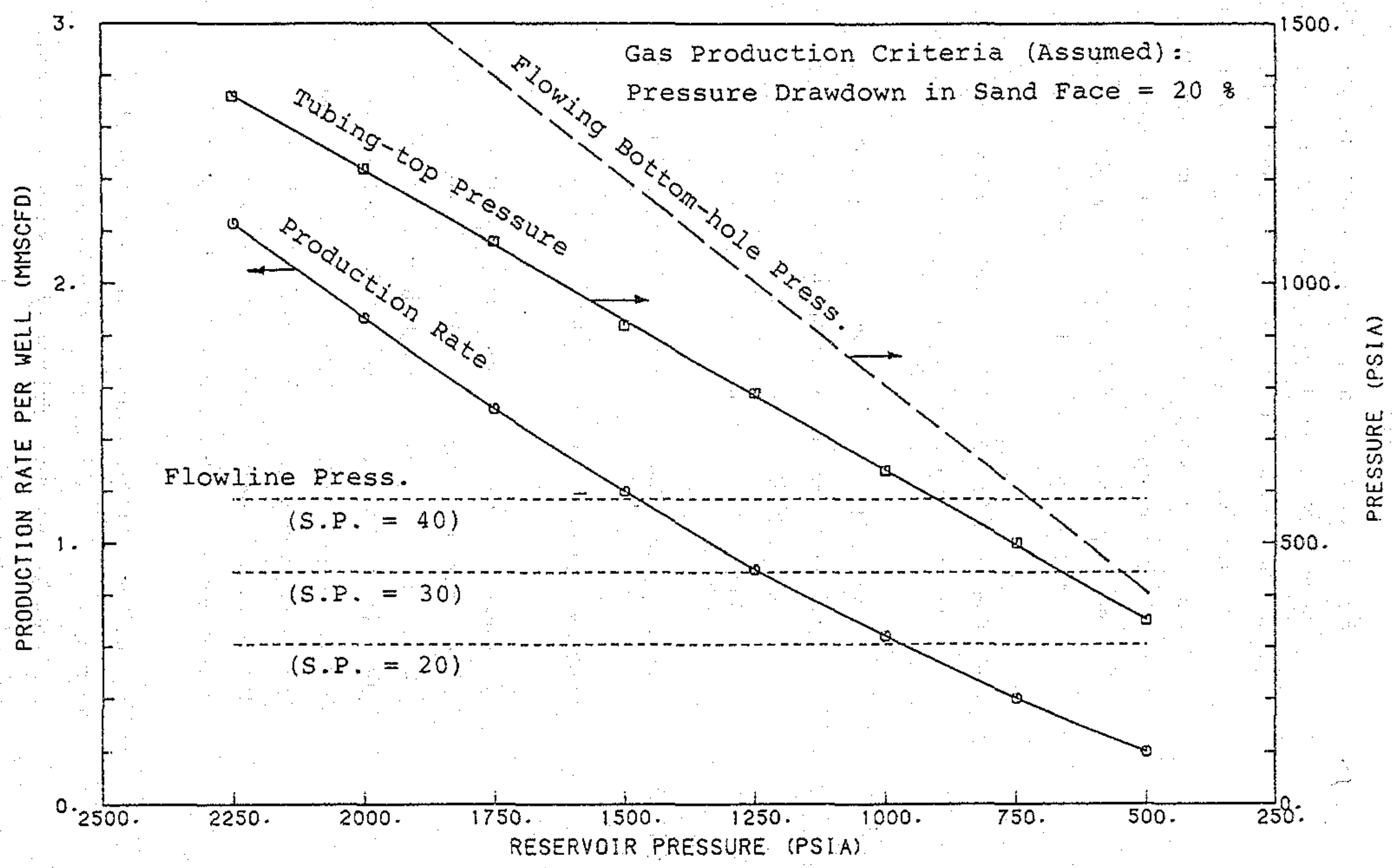


Figure 5-20 Predicted Individual Well Performance

Daily Average		Production Rate per Well			
Field-Total Rate		10th year		20th year	
(10^3 m^3/d)	(MMSCFD)	(10^3 m^3/d)	(MMSCFD)	(10^3 m^3/d)	(MMSCFD)
57	2	56.6	2.0	45.3	1.6
85	3	51.0	1.8	34.0	1.2
113	4	45.3	1.6	22.7	0.8
170	6	34.0	1.2	2.8	0.1

In the range of the daily average field production rate of approximately 57 - 113 x 10^3 m^3/d (2 - 4 MMSCFD), only two or three production wells will be sufficient to maintain the required field-total production rate during the initial and middle stages of the production term. However, in the final stage, the number of the wells may have to be increased up to two - six.

(b) Temperature Profile

The temperature of the production fluid changes due to heat exchange with the outside while the fluid flows from the reservoir to the separator. The temperature change mostly depends on the production rate and the pressure drop at the choke. Therefore, the following three (3) subjects were examined by simplified methods, assuming the well performance shown in Figure 5-20. Then, the temperature profile in the production system was estimated as shown in Figure 5-21.

1) Wellbore temperature change (heat exchange between the formation and the fluid in the well).

2) Temperature drop at the choke (Joule - Thomson effect).

3) Temperature change at the surface flowline (heat exchange between the atmosphere and the flowing fluid in the flowline).

The most typical temperature profile is shown by a solid line in Figure 5-21. Since there are variations in the conditions, the possible range of the temperature profiles due to the variations is shown by broken lines. As per the figure, the fluid temperature at the separator is almost equal to the atmospheric temperature (26.7°C or 80°F in its mean average). This is because the flowline diameter is so large, compared with the gas production rates, that the flow velocity in the pipe is extremely low to allow almost complete heat exchange (heat absorption or heat emission) between the flowing fluid and the atmosphere.

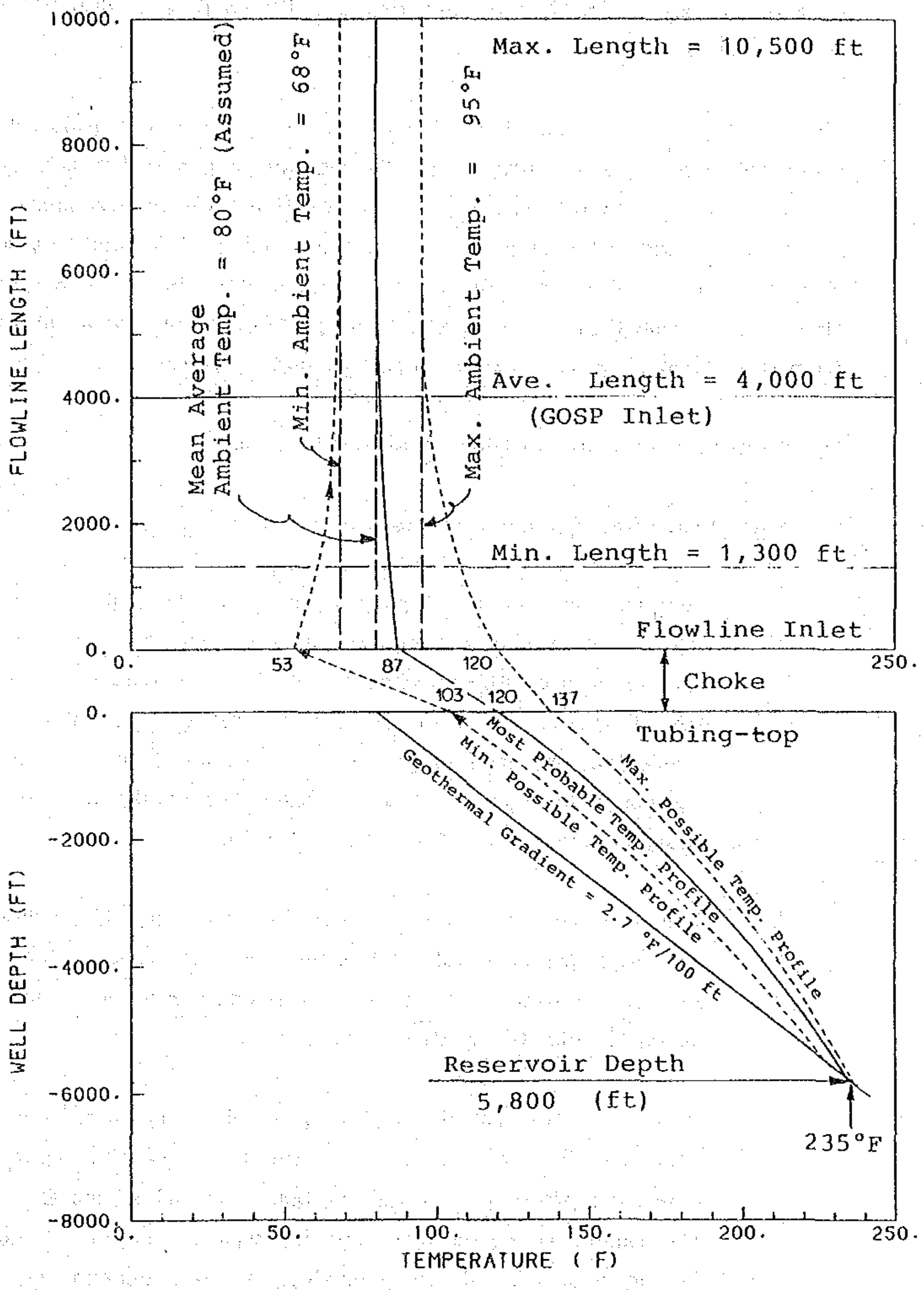


Figure 5-21 Temperature Profile of Production System

5-2-3 *Considerations on Production Facilities and Fluids Delivered to Downstream Plants*

When this project is implemented, the natural gas will be produced in order to supply gas and condensate to the power generation and LPG recovery plants. In this section will be identified the requirements on modification and revamping of the existing production facilities and the delivery conditions of the gas and condensate. In this consideration, it is assumed that no simultaneous production of crude oil etc. from other zones is made to the facilities under discussion.

(1) Considerations on Existing Facility Utilization

(a) Processing Capacity of Separator Trains

Judging from the process flow, the utilization of the existing facilities seems to cause no problems; however, there are no data on the existing separator train capacity except the volume of the high pressure separators (see Table 5-3). Therefore the train capacity was evaluated by the vessel volume assuming that the processing capacity of the HP separator represents the train capacity.

The maximum operation pressure is limited by the mechanical deign pressures of each separator. The production and test separators have different design pressures; about 70 kg/cm^2G and 39 kg/cm^2G respectively.

Based on the result of the production systems analysis described in Sub-section 5-2-2, the range of the separator operation pressure was set at about 20 - 40 kg/cm^2G. For the range, the separator capacity was evaluated as shown in Figure 5-22. Since separator capacity depends on the shape (diameter and length) and internal design etc., a standard type is presumed due to the absence of such information.

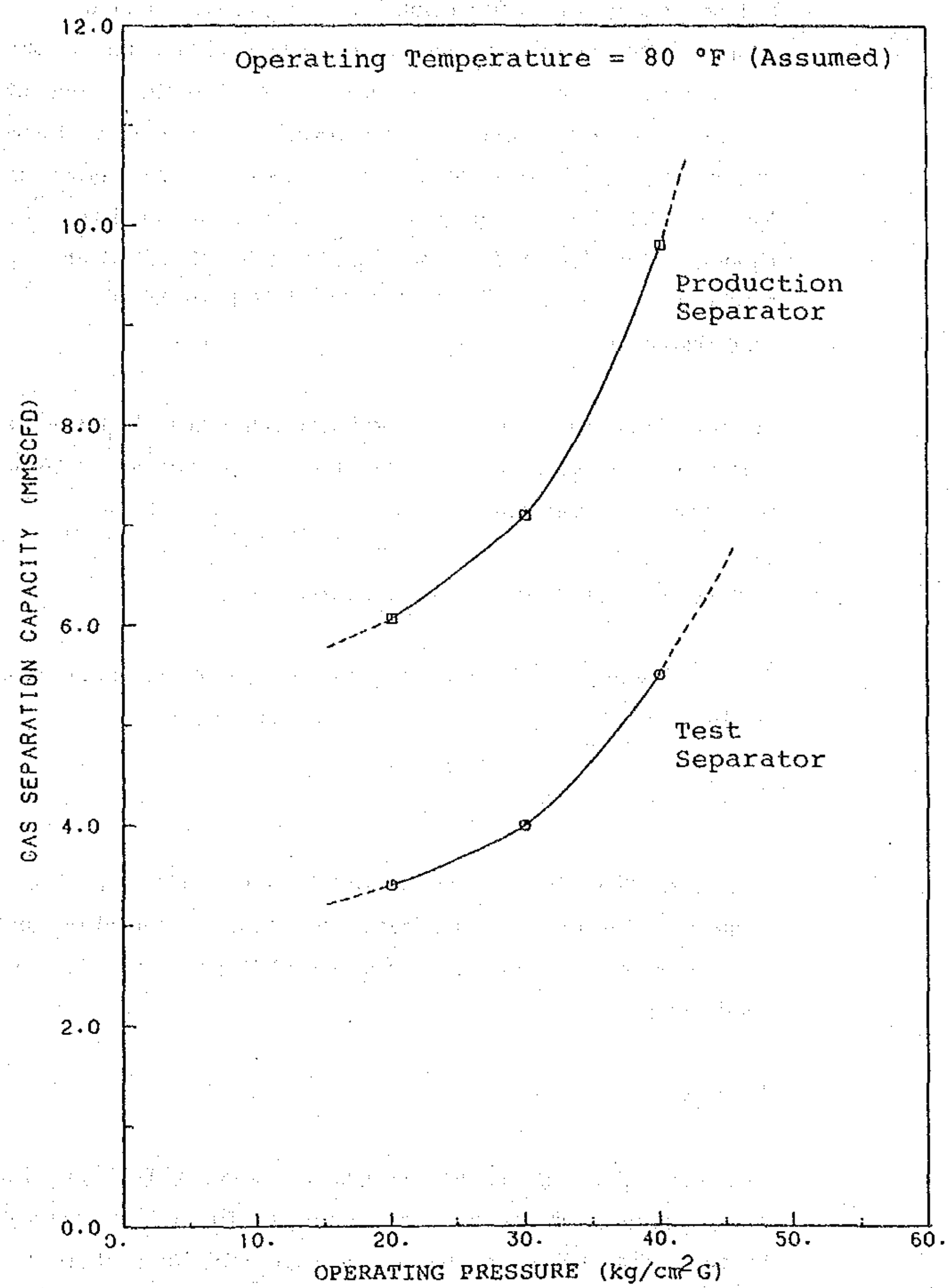


Figure 5-22 Operating Capacity of High Pressure Separators

Unlike crude oil production, the separator capacity is mainly determined by the in-situ gas volume under the operation temperature and pressure, since the liquid amount is comparatively small. Therefore, the capacity turns larger with the increase of the pressure. As shown in Figure 5-22, the capacity gets its lowest at 20 kg/cm^2G, though it is still 156 x 10^3 Nm^3/d (5.5 MMSCFD) in the production train, and 88 x 10^3 Nm^3/d (3.1 MMSCFD) in the test train.

Judging from the above, the existing separator trains have sufficient capacity as long as the gas production rate is not increased drastically.

(b) Requirements on Facility Modification

The following three cases are considered as candidate methods of utilization of the existing facilities.

Case 1:

The case that HP separator gas is sent to the gas pre-treatment facility as fuel gases for power generation and HP separator condensate is fed to LPG recovery plant as feed stock.

Case 2:

The case that HP separator gas is used as fuel gas for power generation as well as feed gas for LPG recovery. In this case, HP separator condensate will be sent to the condensate storage tank after separation of gas at MP and LP separators. MP and LP separator gases will be flared at the flare stack.

Case 3:

The case that separated gases from MP and LP separators in the Case 2 are pressurized and are used as fuel gas for power generation in conjunction with the HP separator gas.

As described in Chapter 6, the method of Case 1 will be optimum scheme for this project. In this case, the following modification work will be required for the existing facilities.

1) The liquid outlet piping of the HP test and production separators are branched, and a pipeline and transfer pumps is installed to deliver the HP condensate to the LPG recovery plant.

2) A pipeline is installed to deliver only the HP gas to the natural gas pre-treatment plant from the gas manifold at the downstream of the HP gas scrubber.

3) A boot is to be provided for condensed water separation at the bottom of HP separator.

(c) Revamping of Production Facilities

To implement this project, the following revamping work is required, in addition to the facility modifications described in (b).

(i) Production Wells

According to PERTAMINA, the following items of work are required for the production wells. This is because, as was described before, some of the production wells are dry-holes and many of the other wells would have troubles such as tubing leakage.

Workover : 8 wells
Recompletion : 11 wells

As is described in Sub-section 5-2-2, only a few wells will be required at the initial stage. Therefore, the initial work could be limited to the minimum required workovers and other work items be left until the necessity arises.

(ii) Gathering Facility and Gas Oil Separation Plant

The following pre-conditioning work is desirable to be carried out for these facilities.

1) Flow lines and plant piping

Pressure test
Corrosion check
Repainting (if necessary)
Pipe internal cleaning (pigging or flushing)

2) Pressure vessels in GOSP

Pressure test
Corrosion check
Internal flushing
Repainting
Function test of instruments and accessories

3) Tanks

Internal cleaning
Corrosion check
Water fill-up test
Repainting, internal coating

4) Oil transmission pumps

The pumps have already been removed. They must be re-installed.

5) Other removed equipment

Basically, all equipment is necessary to be re-installed.

6) Confirmation is needed as for range of measurement and control for all instruments and controllers.

(2) Basic Conditions on Gas and Condensate Delivery

(a) Effect of Separator Operating Condition

The composition of the gas and condensate supplied to downstream plants may vary significantly with the separator operating condition. Tables 5-7, 5-8, and 5-9 summarize the separated gas compositions of HP separation, and the liquid (condensate) compositions at the HP separator outlet corresponding to the operation pressures of the HP separator; 20, 30 and 40 kg/cm^{2}G. As is mentioned in Sub-section 5-2-2, the separator operation temperature is assumed to be 26.7°C (80°F) for the calculation.

The contrast of the HP separator gas compositions is illustrated in Figure 5-23 with varying operation pressures and temperatures. As described in Sub-section 5-2-2, the separator operating temperature will not exceed 26.7°C (80°F) frequently; however, it might be lower. The values at 15.6°C (60°F) are given in the figure as an extreme case. Comparing the accuracy of the reservoir gas composition, the high pressure gas compositions do not show big difference with varying operating pressures and temperatures.

Table 5-7 Composition of Separated Gas and Condensate for 20 kg/cm^2G Separator Pressure (Assumed Gas Composition)

Component		Composition *1		
		Feed	Gas	Liquid
CO_2		6.15	6.52	1.62
C1		51.57	55.17	7.23
C2		14.82	15.30	8.90
C3		15.15	14.34	25.12
iC4		3.48	2.86	11.12
nC4		5.28	4.00	21.05
iC5		2.06	1.13	13.48
nC5		1.38	0.67	10.16
C6+		0.11	0.01	1.33
Mol. Weight		29.80	27.94	52.7
Mols		100.00	92.48	7.5
Water Content	*2	1.4700	0.1850	
	*3	1.47	0.20	

Notes:

*1) *Composition; mol %*
*2) Unit = mols/100 mols-Feed
*3) Unit = mols/100 mols-Stage Separated Gas

Table 5-8 Composition of Separated Gas and Condensate for 30 kg/cm^2G Separator Pressure (Assumed Gas Composition)

Component		Composition *1		
		Feed	Gas	Liquid
CO_2		6.15	6.81	2.44
C1		51.57	58.69	11.54
C2		14.82	15.31	12.06
C3		15.15	12.82	28.23
iC4		3.48	2.25	10.41
nC4		5.28	2.97	18.25
iC5		2.06	0.73	9.55
nC5		1.38	0.41	6.82
C6+		0.11	0.01	0.69
Mol. Weight		29.80	26.53	48.17
Mols		100.00	84.89	15.11
Water Content	*2	1.4700	0.1358	
	*3	1.47	0.16	

Notes:

*1) Composition; mol %
*2) Unit = mols/100 mols-Feed
*3) Unit = mols/100 mols-Stage Separated Gas

Table 5-9 Composition of Separated Gas and Condensate for 40 kg/cm^2G Separator Pressure (Assumed Gas Composition)

Component		Composition *1		
		Feed	Gas	Liquid
CO_2		6.15	6.96	3.20
C1		51.57	61.29	16.14
C2		14.82	14.98	14.25
C3		15.15	11.52	28.38
iC4		3.48	1.89	9.27
nC4		5.28	2.45	15.59
iC5		2.06	0.58	7.46
nC5		1.38	0.33	5.22
C6+		0.11	0.01	0.49
Mol. Weight		29.80	25.66	44.89
Mols		100.00	78.46	21.54
Water Content	*2	1.4700	0.1098	
	*3	1.47	0.14	

Notes:

*1) Composition; mol %
*2) Unit = mols/100 mols-Feed
*3) Unit = mols/100 mols-Stage Separated Gas

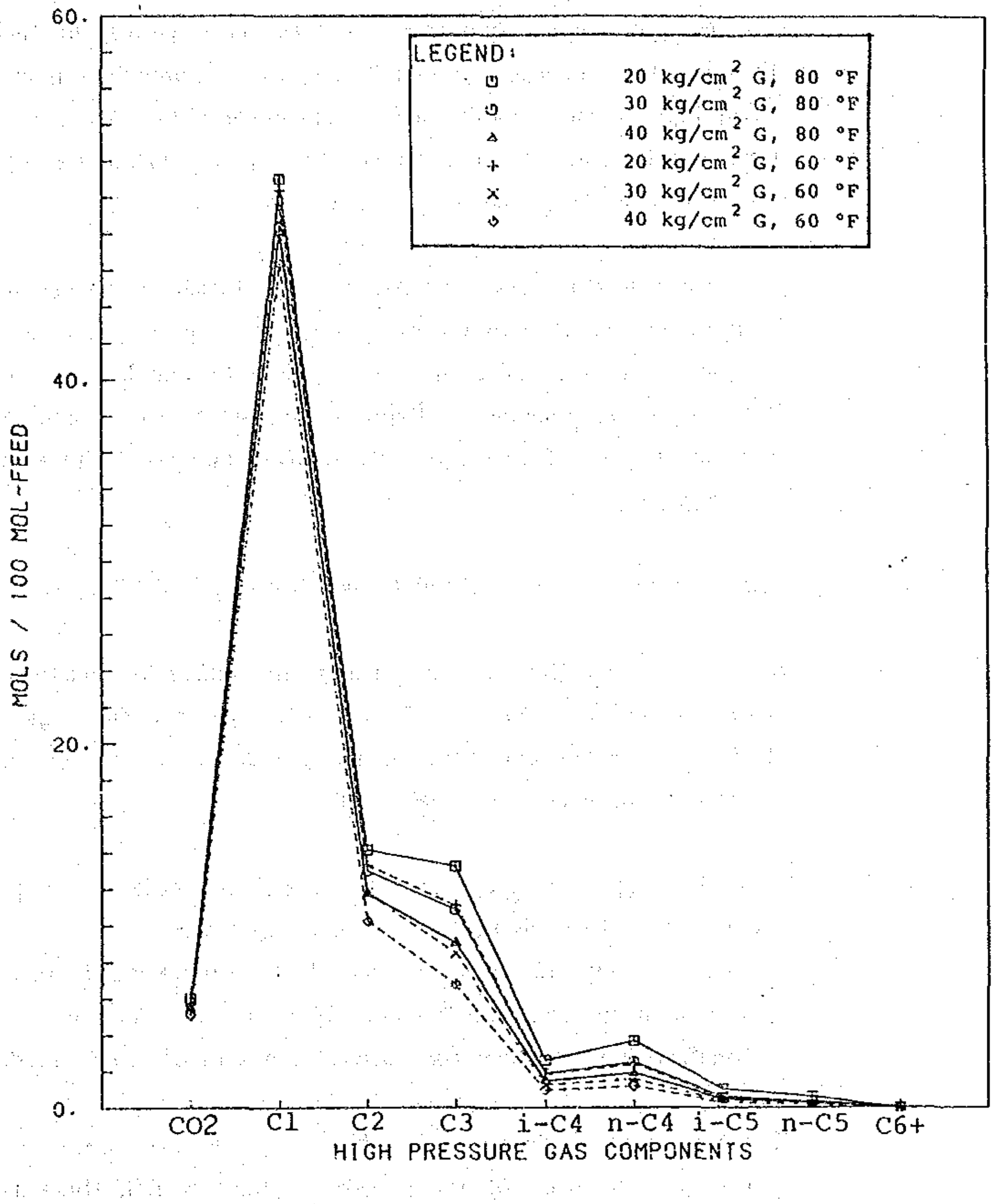


Figure 5-23 Effect of Separator Pressure and Temperature on High Pressure Gas Components

Figure 5-24 shows the volumetric change of the HP separator gas/condensate for 28 x 10^3 m^3/d (1 MMSCFD) reservoir gas. With the increase of the operating pressure, the high pressure gas volume decreases, and the condensates at the HP separator increase. In particular, the volumetric change of the high pressure condensate is remarkable; the condensate volume delivered to the LPG recovery plant varies largely with the separator operating pressure.

The delivery fluid compositions for the different reservoir fluid composition, which was given for reference in Sub-section 5-2-1, are listed in Table 5-10, 5-11 and 5-12 again for a reference purpose. Even in this case, the compositions of delivery fluids show little difference from those in the original case.

(b) Delivery Conditions and Fluid Compositions at GOSP Outlet

After reviewing the following items, the operating pressure of HP separator is set at 25 kg/cm^2G. HP separator gas/condensates ratio and their compositions at the operating conditions are given in Table 5-13.

. Stable supply of gas can be maintained without a gas compressor throughout the entire project life.
. Approximately 10 tons per day LPG production is possible in order to meet the demand in Jambi province.
. Required gas pressure for power plant can be maintained.

However, judging from the number of production wells required for operation of the project (2 to 3 wells), there is possibility that actual composition of the feed gas may be different from those as set in this report. The gas compositions will also vary when production well is changed. Therefore, it is necessary that HP separator has some flexibility in operating pressure in order to maintain supply of constant composition gas to the power plant regardless of variation of feed gas composition.

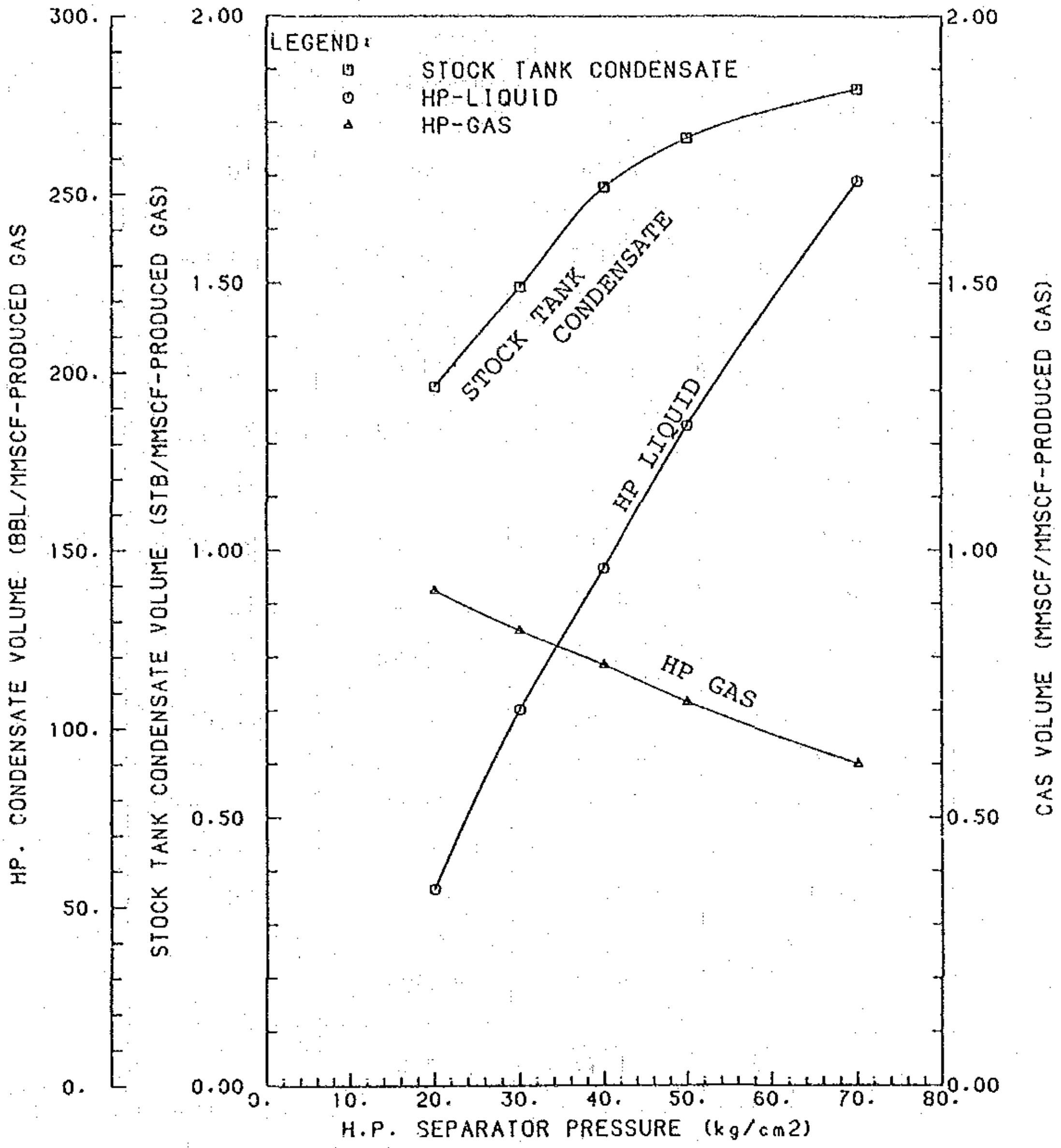


Figure 5-24 Effect of Separator Pressure in Volume of Gas and Condensate

Table 5-10 Composition of Separated Gas and Condensate for 20 kg/cm^2G Separator Pressure (for Reference)

Component		Composition *1		
		Feed	Gas	Liquid
CO_2		6.00	6.89	1.69
C1		48.00	56.37	7.26
C2		14.00	15.08	8.73
C3		15.00	13.29	23.33
iC4		4.00	2.67	10.45
nC4		7.00	4.03	21.43
iC5		3.00	1.04	12.54
nC5		2.00	0.58	8.93
C6+		1.00	0.05	5.64
Mol. Weight		32.18	27.58	54.50
Mols		100.00	82.95	17.05
Water Content	*2	1.4700	0.1659	
	*3	1.47	0.20	

Notes:

*1) Composition; mol %
*2) Unit = mols/100 mols-Feed
*3) Unit = mols/100 mols-Stage Separated Gas

Table 5-11 Composition of Separated Gas and Condensate for 30 kg/cm^{2}G Separator Pressure (for Reference)

Component		Composition *1		
		Feed	Gas	Liquid
CO_2		6.00	7.17	2.53
C1		48.00	60.32	11.59
C2		14.00	14.81	11.61
C3		15.00	11.47	25.43
iC4		4.00	2.07	9.71
nC4		7.00	3.01	18.80
iC5		3.00	0.72	9.73
nC5		2.00	0.40	6.74
C6+		1.00	0.03	3.86
Mol. Weight		32.18	26.13	50.00
Mols		100.00	74.72	25.28
Water Content	*2	1.4700	0.1196	
	*3	1.47	0.16	

Notes:

*1) Composition; mol %
*2) Unit = mols/100 mols-Feed
*3) Unit = mols/100 mols-Stage Separated Gas

Table 5-12 Composition of Separated Gas and Condensate for 40 kg/cm^2G Separator Pressure (for Reference)

Component		Composition *1		
		Feed	Gas	Liquid
CO_2		6.00	7.28	3.30
C1		48.00	63.14	16.15
C2		14.00	14.24	13.50
C3		15.00	10.12	25.26
iC4		4.00	1.74	8.74
nC4		7.00	2.51	16.44
iC5		3.00	0.60	8.05
nC5		2.00	0.33	5.51
C6+		1.00	0.03	3.04
Mol. Weight		32.18	25.25	46.71
Mols		100.00	67.77	32.23
Water Content	*2	1.4700	0.0949	
	*3	1.47	0.14	

Notes:

*1) Composition; mol %
*2) Unit = mols/100 mols-Feed
*3) Unit = mols/100 mols-Stage Separated Gas

Table 5-13 Composition of Separated Gas and Condensate for 25 kg/cm^2G Separator Pressure (Assumed Gas Composition)

Component		Composition *1		
		Feed	Gas	Liquid
CO_2		6.15	6.69	2.17
C1		51.57	57.26	9.30
C2		14.82	15.38	10.66
C3		15.15	13.51	27.35
iC4		3.48	2.49	10.84
nC4		5.28	3.34	19.71
iC5		2.06	0.85	11.08
nC5		1.38	0.49	8.03
C6+		0.11	0.01	0.87
Mol. Weight		29.80	27.07	50.12
Mols		100.00	88.14	11.86
Water Content	*2	1.4700	0.1691	
	*3	1.47	0.19	

Notes:

*1) Composition; mol %
*2) Unit = mols/100 mols-Feed
*3) Unit = mols/100 mols-Stage Separated Gas

5-3 Consideration for Project Implementation

As was mentioned at the beginning of this chapter, the entire study on the natural gas production stands simply on the assumptions in its fundamental part, because of the lack of sufficient data. As a result, the validity of this analysis as a whole, including the compositions of delivered gas and condensate to the power plant and LPG recovery plant, has a strong dependence on the following two points;

1) Possibility of producing crude oil, etc. from other zones than the target ten gas zones simultaneously. In such a case, the composition of the combined fluid at the inlet of the GOSP will differ significantly, and so will do the compositions of the gas and condensate delivered to the downstream plants.

2) The validity of the assumed reservoir characteristics and reservoir fluid composition. When these assumptions deviate significantly from the actual conditions, reliability of the conclusions as a whole will be lost.

These uncertainties probably will not be so serious as to give a fatal influence upon the whole project; however, the parameters developed in this study will not meet the required quality for being used as the plant design data. At the implementation stage of the project, the natural gas production should be re-examined in detail based on more reliable data in order to establish proper parameters and requirements for the plant detail design. Indispensable for the fulfillment of such a study are the understanding and cooperation of PERTAMINA, since the detailed data on the underground resources belong to confidential matter of PERTAMINA.

The following study items should be covered by the additional examination:

1) Establishment of a firm production plan.
2) Evaluation of the reservoir and the performance prediction.
3) Production systems analysis.
4) Finalization of the plant detail design data.

6. PROJECT SCHEME

6-1 Process for Determining Project Scheme

6-1-1 Necessity of Determining Project Scheme

As described in Chapter 1, the object of the Project is to promote economic development by utilizing the unused natural gas in the Sengeti gas field.
The following two plans are considered as an integral part of the project.

(1) To generate electricity using natural gas as fuel and to fulfill an increase of electricity demand being forecast for the region.

(2) To supply LPG to meet LPG demand of Jambi district by recovering LPG fraction contained in the natural gas mentioned above.

In general, natural gases produced from gas reservoirs consist of various components and will be processed to meet the requirements of the various ways they will be consumed. For power generation and LPG recovery contemplated for this project, various methods and processes are used widely in actual applications. Therefore, in order to evaluate and implement the project, it is necessary to determine the optimum combination of processing facilities, capacities of facilities and locations of facility installation, in considering characteristics of feed natural gas and volume of gas reservoirs with relation to demands of electricity and LPG as the final products, in other words, a determination of the project scheme is necessary.

6-1-2 Fundamental Policy in Determining Project Scheme

In determining the project scheme, it should be first decided whether the power generation shall have priority over LPG recovery or not.

For this purpose, listed below are capacities of faciliites and amounts of the natural gas required for power generation and LPG production based on each demand forecast;

	Production Capacity	Required Natural Gas
Power Generation	20 MW	3.54 MMSCFD*
LPG Recovery	10 tons/day	0.15 MMSCFD**

Note:

* Calculated based on low calorific value of 12,580 kcal/ Nm^3, thermal efficiency of 36% and generator efficiency of 96%.

** Figure converting 10 tons per day LPG to gas volume.

As clear from above figures, since the natural gas used for the power generation is overwhelmingly larger in its volume than that of the LPG recovery, the power generation should have priority over the LPG recovery in determining the project scheme.
Therefore, the project scheme should be determined by the following sequence.

(1) Determination of the capacity of power generation and the method of power generation

(2) Determination of the location of the power plant

(3) Determination of a natural gas transmitting method and pretreating facilities

(4) Determination of LPG recovery process, recovery amount and installation site of the LPG recovery plant

6-2 Determination of Power Generation System

6-2-1 Discussion of Scale of Power Generation

According to the electricity demand forecast described in market study, it is anticipated that supply shortage may be taken place for peak load demand in 1994/95 and that the capacity for the power plant contemplated in this project is expected to be in the range of 20 MW. On the other hand, supposing operation period for the power plant will be 20 years in view of mechanical life of a power generation facility, the maximum gas production capacity will be in the range of 5.5 MMSCFD (6,100 Nm^3/hour) from viewpoint of the gas reservoir volume. The amount of power generation by using 5.5 MMSCFD gas, depending on a method of power generation, will be 20 MW to a maximum of 35 MW.

In the light of the two points above, namely, the increase of electricity demand and the limitation of gas supply, a capacity of the power plant is determined at 20 MW.

As mentioned in Chapter 5, the supply amount of feed gas available from Sengeti natural gas reservoirs will be adjusted by chokes installed at well heads. However, volume control by means of a choke is not very easy in actual operation and once set at a certain flow rate, supply volume will not be varied unless conditions of reservoirs change. Therefore, the power plant using a natural gas as fuel should be used as a base load power station, generating a rate of electricity which is hardly changed. Conversely, the existing power plants using diesel oil as fuel which can easily be adjusted to follow load variation shall be used as a load-following power plant, to take on peak loads as they occur.
In case that 20 MW rated power plant is used continuously as a base load power plant, 85% of the rated capacity or 17 MW will be the appropriate amount of power generation. The base load in Jambi district is forecasted to exceed 17 MW in 1995. Therefore, the power plant is decided to generate 17 MW continuously as a base load plant.

6-2-2 Discussion on Power Generation Method

(1) Power Generation System

The following power generation systems are considered for a 20 MW power plant using a natural gas as fuel.

(a) Gas Turbine Generation
(b) Steam Turbine Generation
(c) Dual-Fuel Engine Generation
(d) Spark-Ignited Gas Engine Generation
(e) Gas Turbine-Steam Turbine Combined Cycle Generation

The unit capacity of a spark-ignited gas engine generation is rather small (1 MW range), thus this generation system is not suitable for a 20 MW class power plant. Although a combined cycle generation system is applicable for a 20 MW power plant, the system is very complicated and recently developed for actual application. It is considered too early to introduce this sophisticated system to the Jambi project. Therefore, three generating systems from (a) to (c) above are discussed here in detail.

(2) Economic Comparison

Each generating system has its own merits and demerits and it is not easy to compare them uniformly. Thus, economic comparison is made based on the following premises:

(a) Rated output capacity is 20 MW for all generating systems and power plant does not have a stand-by unit.

(b) Annual generating output is set as 136,000 MWh (8,000 hour operation at 17 MW normal output).

(c) Thermal efficiency is based on low calorific value. In considering uniform operability in actual operation, generating cost is calculated using realistic thermal efficiencies.

(d) Construction cost of power plants is based on standard prices of suppliers.

(e) Capital charge (Depreciation, Interest, Maintenance and Repair Cost, etc.) is defined as a percentage figure of the construction cost. In comparing economics of three generating systems, the capital charge of 25% (Case A) and 12% (Case B) is used.

(f) Price of Diesel Oil (HSD; High Speed Diesel Oil) is set at 0.1201 U.S.Dollars per liter (200 Rp/liter) and its low calorific value is assumed 8,797 kcal/liter.

(g) Three (3) prices of natural gas are used, namely,

Case 1: 2.53 U.S.Dollars per MMBTU
Case 2: 2.10 U.S.Dollars per MMBTU
Case 3: 1.50 U.S.Dollars per MMBTU

Low calorific value is set as 90% of high calorific value.

(h) Amount of fuel oil to be used for dual-fuel engine is set as 10% of rated output of the engine on calorific value basis.

(i) Amount of lubricating oil to be used for gas turbine system and steam turbine system is negligibly small compared with that of the dual-fuel engine system, so is not considered. Dual-fuel engine is assumed to consume lubricating oil at a rate of 1.6 liter per MWh and its price is 0.91 U.S.Dollars per liter (1,501.5 Rp/liter).

Results of calculation of generation cost using above premises are shown in Table 6-1. As it is clear from the Table, the dual-fuel engine generating system is economically superior for all cases over other generating systems, except A-2 and A-3.

Case A is based on a usual city bank interest rate and not considered as an economical source of funds. The condition that this project is implemented is considered to be the condition close to Case B, which is based on institutional loan or bi-lateral loan condition.
In view of the above results, the dual-fuel engine generation system is considered the most appropriate from economical viewpoint.

Table 6-1 Economic Comparison of Power Generating System

(Unit: US$/year)

	Dual-Fuel Engine	Gas Turbine	Steam Turbine
Engine Capacity	5 MW x 4	10 MW x 2	20 MW x 1
Normal Output	17 MW	17 MW	17 MW
Thermal Efficiency	36%	20%	25%
Construction Cost (US$)	27,500,000	18,000,000	28,500,000
Capital Cost			
Case-A @ 25%	6,875,000	4,500,000	7,125,000
Case-B @ 12%	3,300,000	2,160,000	3,420,000
Fuel Cost			
Gas			
Case-1	3,197,876	6,523,666	5,218,933
Case-2	2,654,363	5,414,900	4,331,921
Case-3	1,895,974	3,867,786	3,094,229
Fuel Oil (Pilot)	521,824		
Lube Oil	198,016		
Annual Cost			
A-1	* 10,792,716	11,023,666	12,343,933
A-2	10,249,203	* 9,914,900	11,456,921
A-3	9,490,814	* 8,367,786	10,219,229
B-1	* 7,217,716	8,683,666	8,638,933
B-2	* 6,674,203	7,574,900	7,751,921
B-3	* 5,915,814	6,027,786	6,514,229

Note: * Shows least expense method for each case

(3) Characteristics of Each Generating System

(a) Gas Turbine Generation System

As it is clear from the comparison of Table 6-1, the characteristic of gas turbine generation system is its low construction cost. Therefore, gas turbine generation is favourable when fuel cost is less expensive and capital cost such as interest is high.

Merits of a gas turbine generation system are the light weight of the machine and its quick start up. And also due to its high revolution-per-minute, governor response is fast and stable compared with the reciprocating machine. Furthermore, the gas turbine generation system does not need cooling water nor radiator. Using those characteristics of the gas turbine generator, the system is applied where soil conditions are poor and access is difficult, or to cope with peak load or as a temporary power station for construction purposes. Thermal efficiency based on low calorific value is set at 20% in the calculation of economic comparison, although a gas turbine of higher thermal efficiency is available. High level skills are required for maintenance and overhaul work.

Because there is no maintenance shop around Jambi City to repair gas turbines and there are locations where soil conditions are strong enough for installation of dual-fuel engines in Jambi district, and transportation of dual-fuel engines is feasible in the district, there is no necessity to select a gas turbine generating system for this project.

(b) Steam Turbine Generation System

20 MW output is rather small in scale for a steam turbine generating system. According to selection of steam conditions, thermal efficiency of steam turbine generator can be increased and some of recent large thermal power plants achieve an efficiency higher than 40% on a high calorific

value basis. The thermal efficiency adopted for this project is 25% on low calorific value basis, which is low compared with the efficiency of recent large thermal power plants. This is because there is no operating experience of steam power plant in the vicinity of Jambi City and steam conditions are set at 40 kg/cm^2G, 450°C considering easy operation of the plant.

Output of steam turbine plant is very much influenced by the temperature of cooling water for the condenser, and its requirement for a large amount of cooling water is considered a demerit. On the other hand, it can use various kinds of fuels including solid fuel such as coal and wooden chips. This is a merit of the steam power plant. Further, in case of steam power plant, it is rather easy to control air pollution such as nitrogen oxides and also it is easy to control its level of noise compared with that of diesel engine or gas turbine. However, since environmental control is not so strict in Jambi City, there is no necessity to adopt steam power plant for the project.

(c) Dual-Fuel Engine Generation System

Demerits of the dual-fuel engine generating system are that it generates a high level of noise, needs liquid fuel as pilot fuel, and its nitrogen oxides emission is difficult to control. However, these demerits will not become obstacles in the practical sense.

The mechanism of dual-fuel engine is identical to that of diesel engine and operating know-how and maintenance technique for the diesel engine can be utilized for dual-fuel engine.

The existing power plants in Jambi City are of all diesel engine generation; consequently it is advantageous to utilize the technology of operators of these diesel engine plants in the operation and maintenance of a dual-fuel engine power plant.

Table 6-2 shows these comparison.

Table 6-2 Comparison of Prime Movers

	Dual-Fuel Engine	Gas Turbine	Steam Turbine
Output	Almost no influence by atmospheric temperature Influenced by fuel (Knocking limit)	Influenced by atmospheric temperature Not influenced by fuel	Influenced by cooling water temperature Not influenced by fuel
Fuel to be used	Gas & Liquid fuel (requested dual)	Gas or liquid fuel	Gas, liquid fuel and/or coal
Cooling water	Radiator or cooling tower Relatively small amount is required.	Not required	Relatively large amount is required.
Installing weight	Relatively heavy	Light	Moderate
Operation & maintenance	Almost same as diesel engine, thus easy for local operator	Special facility for overhauling is required.	Special knowledge is required for local operator
Environmental influence	Relatively high noise level and high NOx	Relatively high noise level	Exhaust gas emission control is relatively easy.

(4) Conclusion

Since the results of economic comparison may easily be changed due to changes of the economic situation which have great influence on capital cost and fuel cost, it is not adequate to draw a conclusion merely by the results of the economic comparison table. However, in addition to the economical advantage, the dual-fuel engine generation system is suitable for this project not only from a technical viewpoint, but also from the viewpoint of maintenance and the following discussions are developed based on the dual-fuel engine system.

6-3 Discussion of Power Plant Site

Comparison is made among the following three candidate locations selected during the field survey.

a) Sengeti
b) A location between Setiti and Jambi City
c) Payo Selincah, adjacent to the existing power plant

6-3-1 Policy in Selecting Site for Power Plant

In case that electricity is generated using natural gas, it is necessary to install pipeline from gas reservoirs to a power plant and to construct electricity transmission and distribution lines. So, in selecting site for a power plant, an overall assessment is required not only from a viewpoint of power plant construction but also from a viewpoint of pipeline installation and contruction of transmission and distribution lines. For this project, a dual-fuel engine generating system with a 20 MW rated output is selected as described in Section 6-2. Therefore, at first technical evaluation is carried out in this section on a pipeline system and a transmission and distribution system to select the optimum route of pipeline and transmission line among those three alternatives. Then, construction cost and energy loss due to transmission will be calculated for each case, and finally the optimum site location for power plant will be selected.

Although there might be some difference in the construction cost of the power plant for each case, they are negligible compared with that of construction costs for pipeline and transmission line, therefore, are not considered for the evaluation.

6-3-2 Discussion on Pipeline System

In case that the power plant is constructed either at Setiti or at Payo Selincah, it is necessary to install a gas pipeline to transfer gas from the gas field to the power plant. The pipeline size required to transfer gas of 3.2 MMSCFD (3,500 Nm^3/h) sufficient to generate 20 MW electricity is 6 inch in diameter (refer to Chapter 9). In the case of Setiti, the pipeline would pass along the existing oil pipeline from Sengeti to Setiti on hilly areas. On the other hand, in the case of Payo Selincah, the pipeline would pass through swampy areas of the northern side of the Batang Hari river and conditions for constructing a pipeline differ significantly.
In this study, considering difficulties of construction, the unit construction costs are established as follows:

- Between Sengeti and Setiti: 8 U.S.Dollars/inch-meter
- Between Sengeti and Payo Selincah: 12 U.S.Dollars/inch-meter

Pipeline lengths from Sengeti gas field to Setiti and to Payo Selincah are 15 kilometers and 20 kilometers, respectively; the construction costs for both cases are as follows:

- Between Sengeti and Setiti
  8 US$/inch-meter x 6 inch x 15 km = 720,000 U.S.Dollars
  = 1,198,800,000 Rp
- Between Sengeti and Payo Selincah
  12 US$/inch-meter x 6 inch x 20 km = 1,440,000 U.S.Dollars
  = 2,397,600,000 Rp

6-3-3 Discussion on Transmission and Distribution System

In the discussion of a transmission and distribution system, it is necessary to formulate a transmission and distribution plan to meet electricity demand by assuming a distribution of demand in the target area. In assuming a demand distribution, since the each category of consumers shows a different demand pattern, it is necessary to assume the demand for each category.
In this project, the electricity demand for the industrial sector and other sector are assumed against each time range based on the development plan of Jambi City and the expansion program of electrification planned by PLN.
Then a future plan will be developed on the transmission and distribution system based on the assumed demand pattern.
The other sector use is the total of household use, public use and commercial use because daily changes in their electricity demands are almost same and similar demand patterns are observed. Time range is defined as follows:

- Morning time : 0:00 hours to 7:00 hours
- Day time : 7:00 hours to 17:00 hours
- Night time : 17:00 hours to 24:00 hours

(1) Assumption of Electricity Demand Distribution

Table 6-3 shows the estimated electricity demand of each consumer sector for 1994 to 1999 based on the Market Study described in Chapter 4, and Table 6-4 shows increases of electricity demand for industrial use and other use until 1999 based on figures of 1994. Although it is difficult to make an accurate assumption of electricity demand for the morning time, day time and night time ranges, the electricity demand for each time range is forecasted on the assumption that a pattern of daily change will remain unchanged in the future, as shown in Figure 6-1.

Electricity demand for industrial use in Jambi district will be maximum during day time and electricity consumption during morning and night time will be somewhat limited to the demand of a few factories running on a 24-hour basis and the demand for lighting during night time. Therefore, it is assumed in this study that 80% of the industrial demand will be consumed during day time (from 7:00 hours to 17:00 hours) and 10% each in the morning and in the night time. The electricity consumption by the other sector, consisting of demands for household use and street lighting, is assumed as 32%, 18% and 50% for morning time, day time and night time, respectively.

Since time of one year is 8,766 hours (considering leap year), electricity demands for each sector are calculated as shown in Table 6-5, based on morning time of 2,556.75 hours, day time of 3,652.5 hours and night time of 2,556.75 hours. The figures in this Table do not represent an increase of peak load, but represent an incremental increase of average value of load.

Table 6-3 Electricity Demand Forecast For Each Sector (1994 through 1999)

(Unit: GWh/year)

F.Year	1994 (Base)	1995	1996	1997	1998	1999
Residential	46.5	50.4	54.7	59.3	64.3	69.8
Commercial	9.1	9.6	10.1	10.6	11.1	11.7
Public	9.2	9.7	10.3	10.9	11.6	12.2
Industry	67.1	75.6	84.5	94.0	104.0	110.3
Total	131.9	145.3	159.6	174.8	191.0	204.0

Table 6-4 Electricity Demand Increase For Each Sector Based on Figures of 1994

(Unit: GWh/year)

F.Year	1995	1996	1997	1998	1999
Industry	8.5	17.4	26.9	36.9	43.2
Others	4.9	10.3	16.0	22.2	28.9
Total	13.4	27.7	42.9	59.1	72.1

Table 6-5 Electricity Demand Increase for Each Time Range

(Unit: kW)

F.Year	1995	1996	1997	1998	1999
Industry					
Morning	332	681	1,052	1,443	1,690
Day	1,862	3,811	5,892	8,082	9,462
Night	332	681	1,052	1,443	1,690
Others					
Morning	613	1,289	2,003	2,779	3,617
Day	241	508	789	1,094	1,424
Night	958	2,014	3,129	4,341	5,652
Total					
Morning	946	1,970	3,055	4,222	5,307
Day	2,103	4,319	6,680	9,176	10,886
Night	1,291	2,695	4,181	5,785	7,341

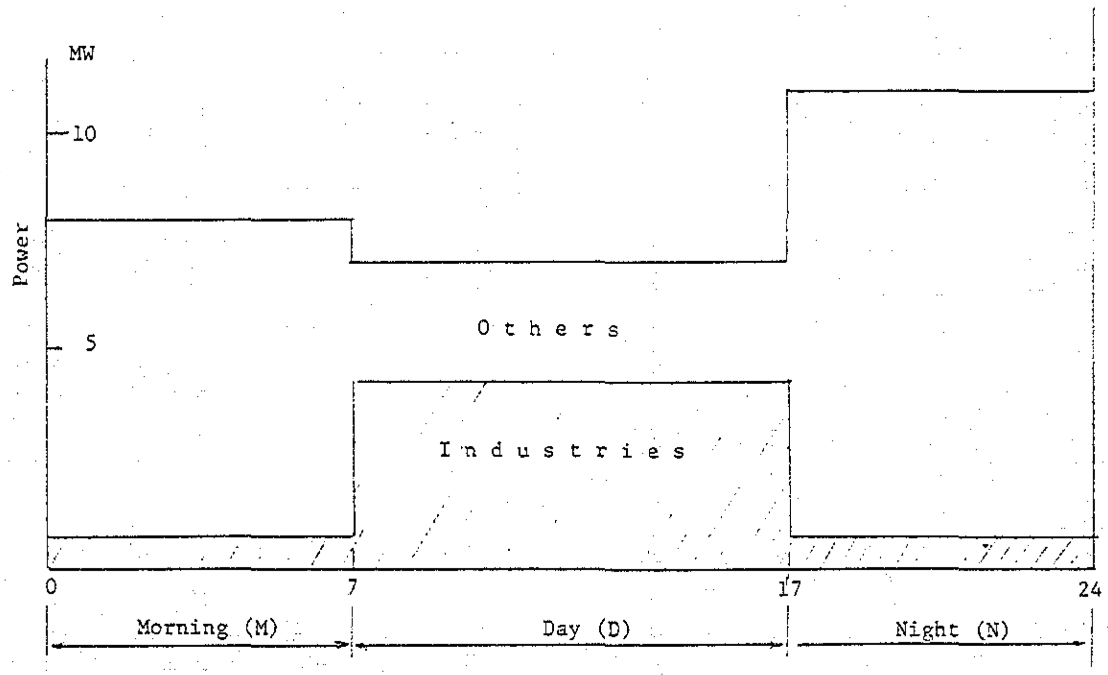


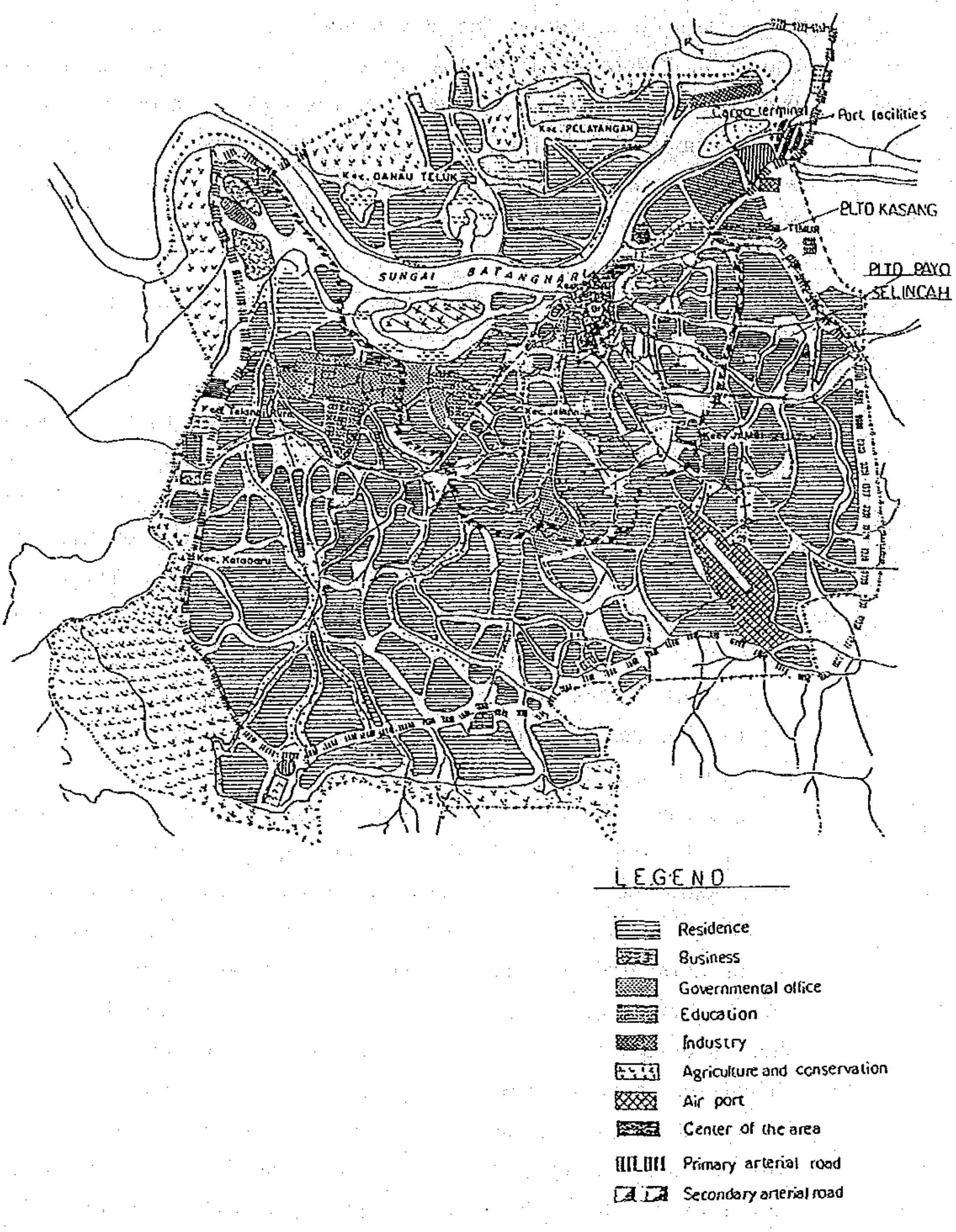
Figure 6-1 Daily Fluctuation of Electricity Demand

It is assumed based on the development plan of Jambi City as shown in Figure 6-2 that the increase of demand for industrial use will occur in the east region of the city. Demand for other use is assumed for the west and the south-west of the city in accordance with electrification program prepared by PLN (refer to Figure 6-3).

(2) Conceptual Plan and Construction Cost of Transmission and Distribution Lines

As mentioned above, the industrial zone is clustered in the east part of Jambi City. Electricity for the industrial use will be sent to the east region and electricity for other use will be transmitted to the west and the south-west regions of the city where public and residential areas are expanding. Therefore, the transmission line routes for Case A (power plant is built in Sengeti), Case B (power plant in Setiti) and Case C (power plant in Payo Selincah) are as shown in Figure 6-4, Figure 6-5 and Figure 6-6, respectively. Figure 6-7 illustrates models of Case A, Case B and Case C.
Once a route for the transmission line is determined, the length of the tie line can be decided. If average length of distribution lines and amount of electricity distributed are assumed to be equal for all three cases and the numbers of distribution lines to be extended are also considered the same regardless of the location of power plant site, construction cost for distribution lines and loss of energy due to transmission become the same; therefore, these can be eliminated from comparison calculation. Table 6-6 shows type, size and length of tie line for comparison, where, UCG-400 represents underground copper cable with a 400 mm^2 cross section, UCG-325, the same cable with a 325 mm^2 cross section and SC-400, a submerged copper cable with a 400 mm^2 cross section.

Unit construction costs of transmission lines are defined as follows, which include all construction cost elements on averaged basis.



__Quoted from Plan of Municipal office_

Figure 6-2 Utilization Plan of City Area

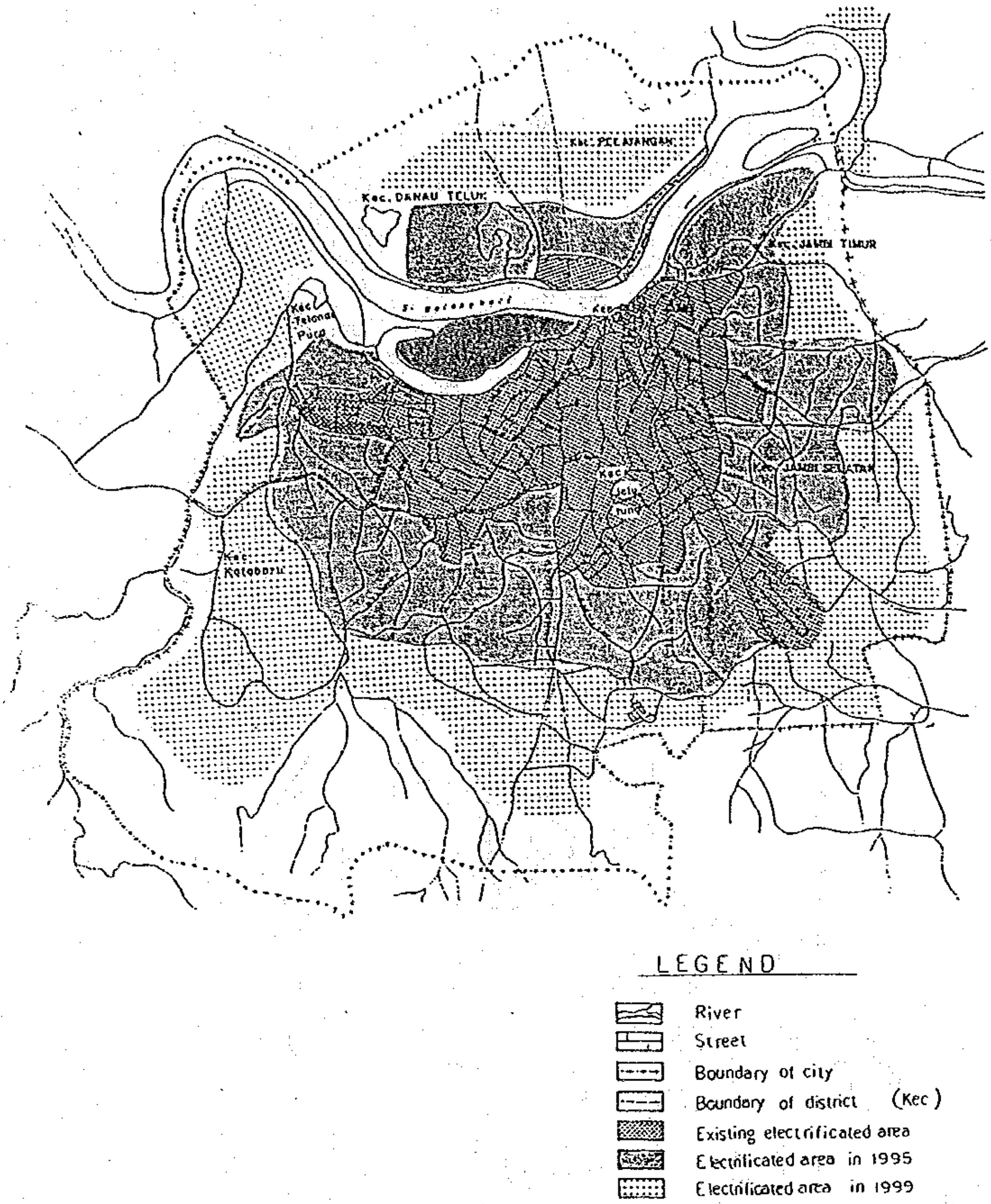


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Figure 6-3 Electrification of Jambi City

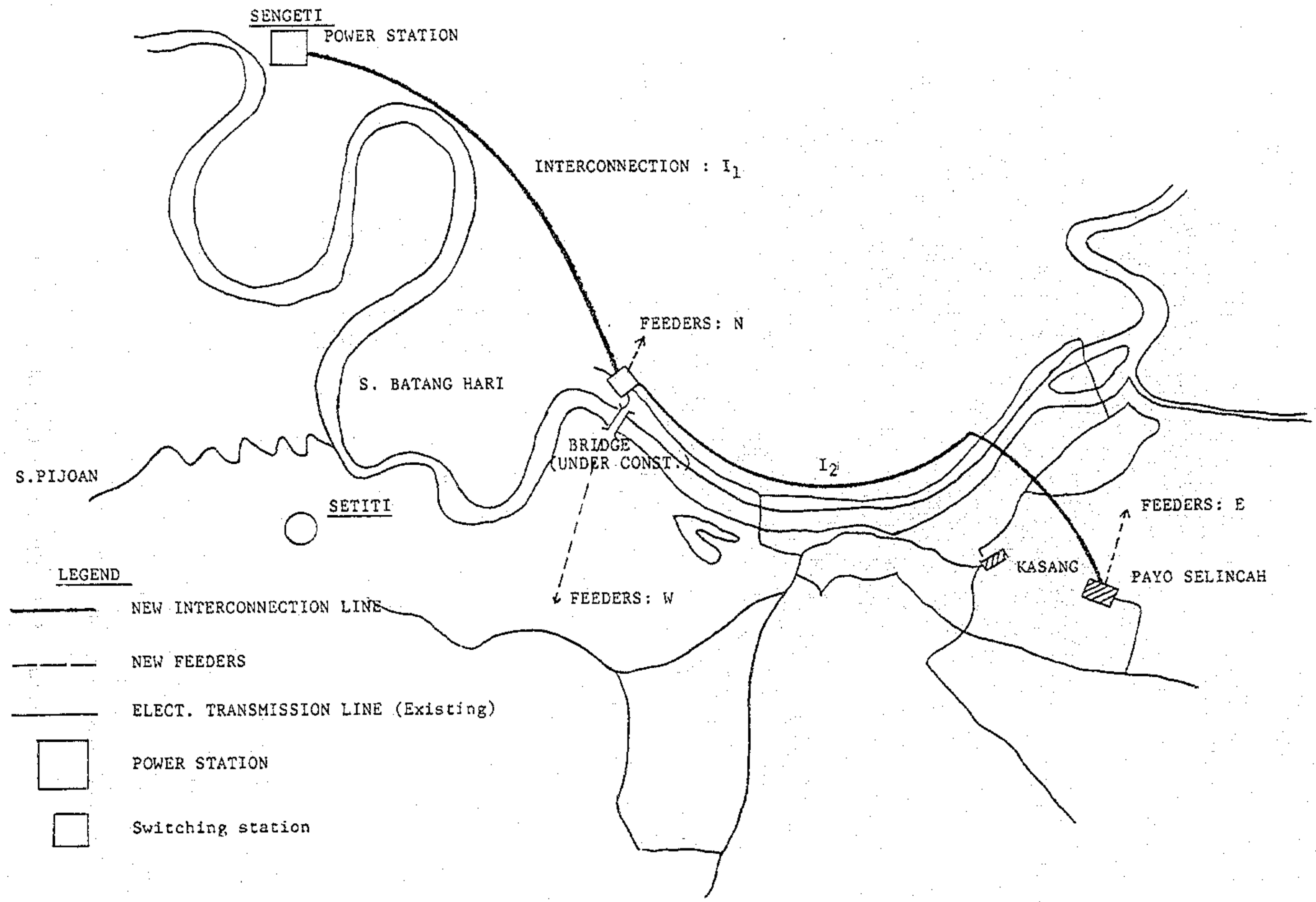

Figure 6-4 Transmission Line Case A

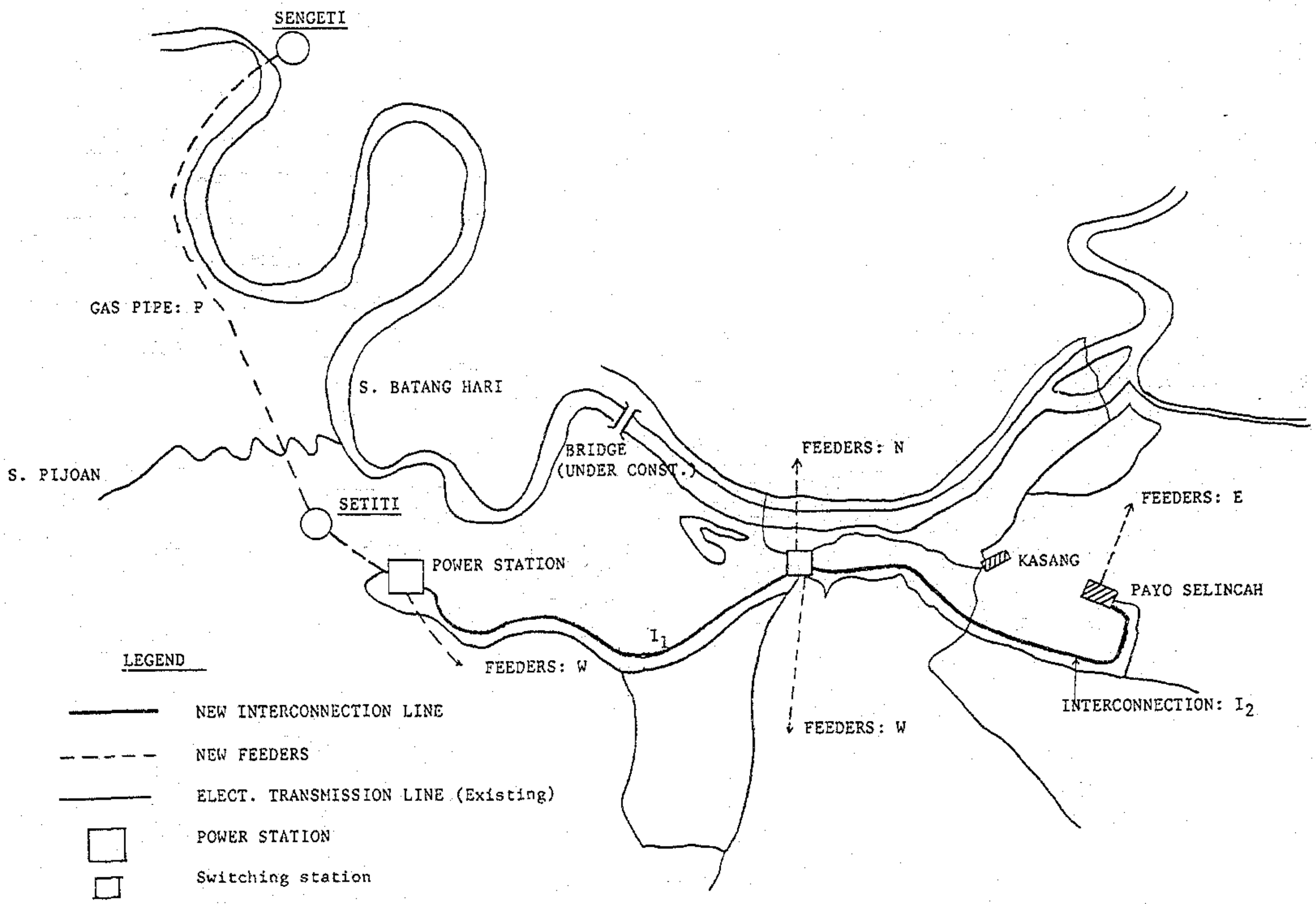

Figure 6-5 Transmission Line Case B

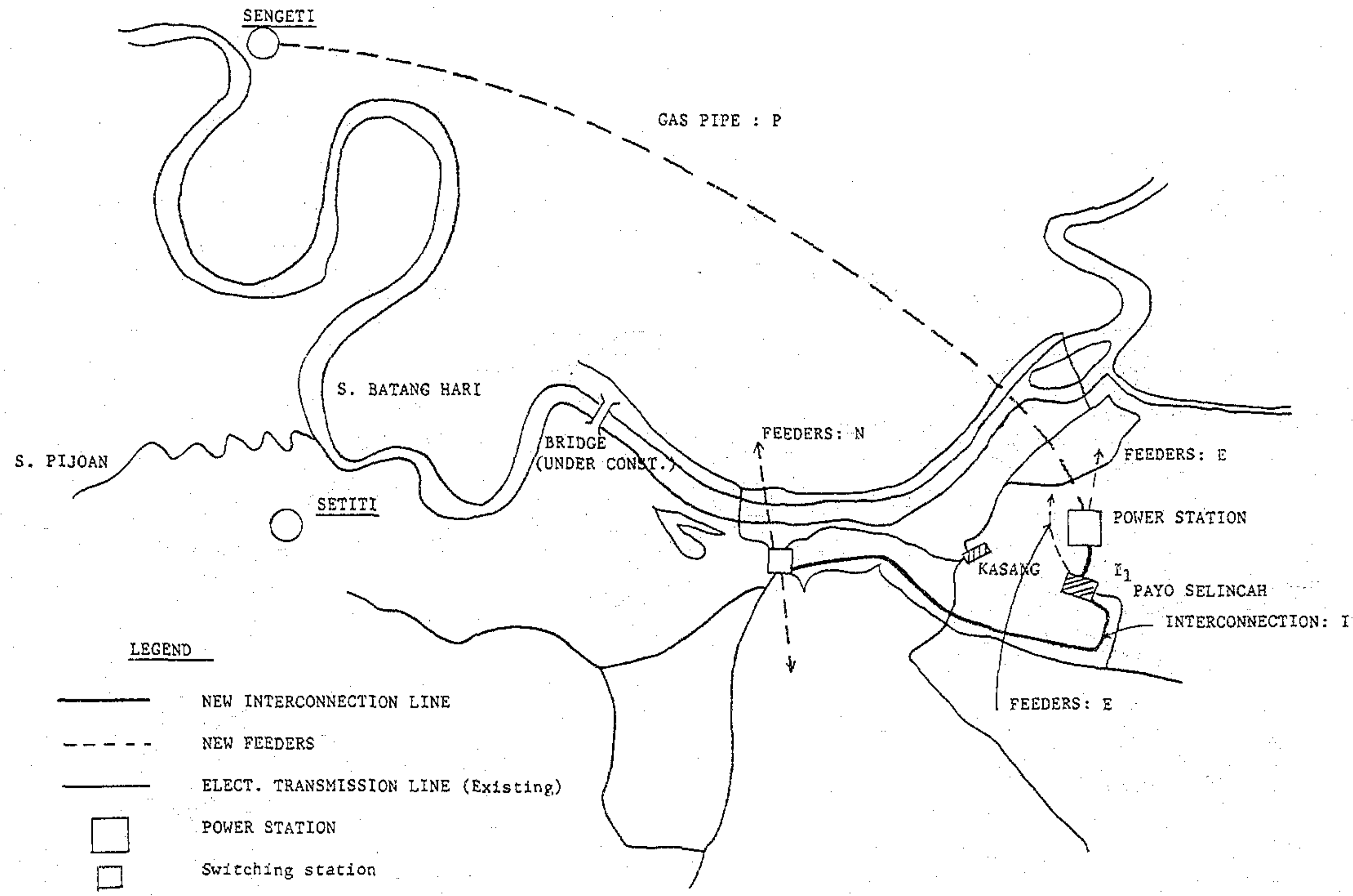

Figure 6-6 Transmission Line Case C

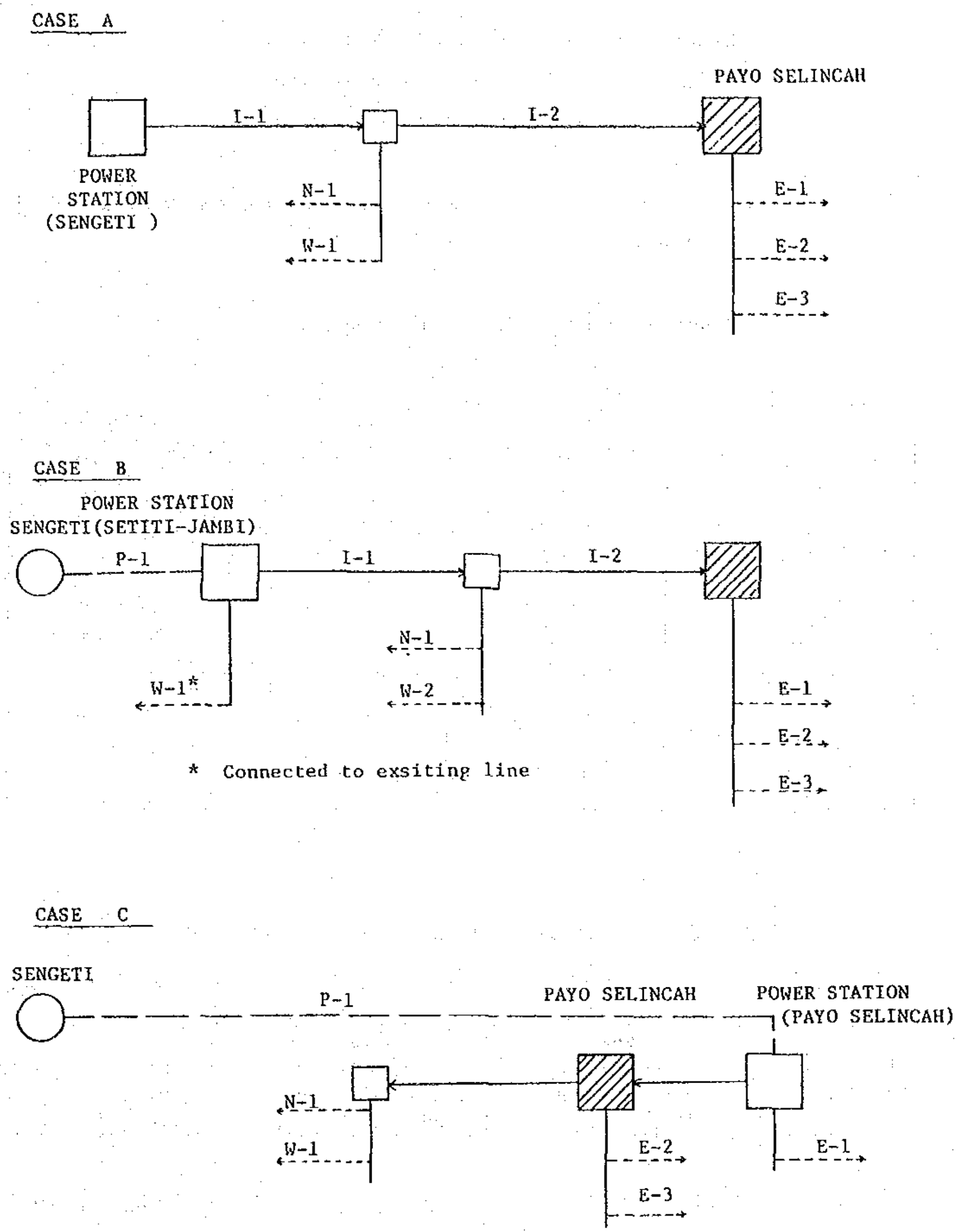

Figure 6-7 Transmission Line Models

- UCG-400 : 85,000,000 Rp/kilometer
- UCG-325 : 80,000,000 Rp/kilometer
- SC-400 : 160,000,000 Rp/kilometer

Construction costs of transmission lines are calculated and shown in Table 6-7.

Table 6-6 Type, Size and Length of Cable

	Case A (Sengeti)	Case B (Setiti)	Case C (Payo Selincah)
Cable			
Type	UCG-400	UCG-400	UCG-325
Length	27.13 km	19.8 km	0.3 km
Type	SC-400	-	UCG-400
Length	0.87 km	-	9.0 km
1-1	12.0 km	10.8 km	0.3 km
1-2	16.0 km	9.0 km	9.0 km

Table 6-7 Comparison of Transmission Line Cost

(Unit: Thousand Rp)

Unit Cost (Rp/km)	Case A (Sengeti)		Case B (Setiti)		Case C (Payo Selincah)	
	Length (km)	Cost	Length (km)	Cost	Length (km)	Cost
UCG-400 85,000	27.13	2,306,050	19.80	1,683,000	9.00	765,000
UCG-325 80,000	-	-	-	-	0.30	24,000
SC-400 160,000	0.87	139,200	-	-	-	-
Total	28.00	2,445,250	19.80	1,683,000	9.30	789,000

(3) Transmission Loss

As explained previously, the interconnection lines are the basis of comparison for energy loss, and resistance value of cable will be decided upon selecting a type of cable for the interconnection lines.
Resistance value of the transmission cables are as follows:

- UCG-400 : 0.04402 per kilometer
- SC-400 : 0.04402 per kilometer
- UCG-325 : 0.03577 per kilometer

In evaluating transmission energy loss, it is necessary to calculate the amount of electricity to be transmitted for each time range, on a year by year basis. Amounts of transmitted electricity are as follows:

Case A:
If a power plant is built at Sengeti, all of the 17 MW of electricity generated will be sent through I-1 to a switching station near the bridge now under construction. Increase of electricity demand required for industrial zone in the east side of the city would be sent through I-2 to Payo Selincah.

Case B:
If a power plant is built is Setiti, 2 MW of the 17 MW of electricity generated will be sent to residential and cultural zones in the west region of the city through W-1 and W-2.
The rest of 15 MW will be sent to the central part of the city through I-1, where electricity is distributed for residential and public uses. To the area of the east of the city to cover increased amount of demand in the industrial zone, electricity will be sent by the similar manner to Case A.

Case C:

If a power plant is built in Payo Selincah, 14 MW of the 17 MW of electricity generated will be sent to the existing Payo Selincah power plant via I-1, and 3 MW will be sent through E-1, E-2 and E-3 to cover the demand in neibouring area. Some of the electricity will be sent to the central part of the city through I-2 to meet an increasing demand in that area.

Table 6-8 shows the amount of electricity to be transmitted for each case as explained above. There are some difference of load impedance on each transmission line in addition to variation of a power factor during transmission. A power factor for this calculation is set at 0.8.

Transmission loss P_L (kW) is calculated by the following formula for a transmission voltage of 20 kV, where R is resistance and P (kW) is transmission power:

$$P_L = \{P/(20x0.8)\}^2 \text{ x } R/1,000$$

Transmission losses calculated using the results of Table 6-8 are shown in Table 6-9.

Table 6-8 Electricity Transmitted through Tie Lines

(Unit: kW)

F.Year		1995	1996	1997	1998	1999
Case-A						
I-1	M	17,000	17,000	17,000	17,000	17,000
	D	17,000	17,000	17,000	17,000	17,000
	N	17,000	17,000	17,000	17,000	17,000
I-2	M	332	681	1,052	1,443	1,690
	D	1,862	3,811	5,892	8,082	9,462
	N	332	681	1,052	1,443	1,690
Case-B						
I-1	M	15,000	15,000	15,000	15,000	15,000
	D	15,000	15,000	15,000	15,000	15,000
	N	15,000	15,000	15,000	15,000	15,000
I-2	M	332	681	1,052	1,443	1,690
	D	1,862	3,811	5,892	8,082	9,462
	N	332	681	1,052	1,443	1,690
Case-C						
I-1	M	14,000	14,000	14,000	14,000	14,000
	D	14,000	14,000	14,000	14,000	14,000
	N	14,000	14,000	14,000	14,000	14,000
I-2	M	613	1,289	2,003	2,779	3,617
	D	241	508	789	1,094	1,424
	N	958	2,014	3,129	4,341	5,652

Table 6-9 Results of Calculation of Transmission Loss through Tie Lines

(Unit: kW)

F.Year		1995	1996	1997	1998	1999	
Case-A							
I-1	M	596	596	596	596	596	R = 0.5282 Ω
	D	596	596	596	596	596	
	N	596	596	596	596	596	
I-2	M	0	1	3	6	8	R = 0.7052 Ω
	D	10	40	96	180	247	
	N	0	1	3	6	8	
Case-B							
I-1	M	418	418	418	418	418	R = 0.4754 Ω
	D	418	418	418	418	418	
	N	418	418	418	418	418	
I-2	M	0	1	2	3	4	R = 0.3962 Ω
	D	5	22	54	101	139	
	N	0	1	2	3	4	
Case-C							
I-1	M	8	8	8	8	8	R = 0.01073 Ω
	D	8	8	8	8	8	
	N	8	8	8	8	8	
I-2	M	1	3	6	12	20	R = 0.3962 Ω
	D	0	0	1	2	3	
	N	1	6	15	29	49	
Total:							
Case-A							
	M	597	598	599	602	604	
	D	606	636	692	776	843	
	N	597	598	599	602	604	
Case-B							
	M	418	419	420	421	422	
	D	423	440	472	519	556	
	N	418	419	420	421	422	
Case-C							
	M	9	11	14	20	28	
	D	8	9	9	10	11	
	N	10	14	23	37	58	

Table 6-10 shows the amount of loss of transmission for each year based on 100 Rp per kWh of electricity cost and time range shown in Table 6-9. Also indicated in Table 6-10 are amounts of transmission loss for each year discounted to 1995 present value, using present value coefficient of 12% which is a standard value adopted by PLN for this sort of calculation.

Table 6-10 Results of Calculation of Transmission Loss through Tie Lines

F. Year	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005
Power Loss (MWh)											
Case-A	5,263	5,380	5,592	5,914	6,168	6,168	6,168	6,168	6,168	6,168	6,168
Case-B	3,683	3,748	3,868	4,048	4,191	4,191	4,191	4,191	4,191	4,191	4,191
Case-C	77	96	130	184	262	262	262	262	262	262	262
Cost of Loss (MMRp)											
Case-A	526.3	538.0	559.2	591.4	616.8	616.8	616.8	616.8	616.8	616.8	616.8
Case-B	368.3	374.8	386.8	404.8	419.1	419.1	419.1	419.1	419.1	419.1	419.1
Case-C	7.7	9.6	13.0	18.4	26.2	26.2	26.2	26.2	26.2	26.2	26.2
Worth of Cost in the year 1995 (MMRp)											
- Rate -	1.0000	0.8929	0.7972	0.7118	0.6355	0.5674	0.5066	0.4523	0.4039	0.3606	0.3220
Case-A	526.3	480.3	445.8	420.9	392.0	350.0	312.5	279.0	249.1	222.4	198.6
Case-B	368.3	334.7	308.3	288.2	266.4	237.8	212.3	189.6	169.3	151.1	135.0
Case-C	7.7	8.6	10.4	13.1	16.6	14.8	13.3	11.8	10.6	9.4	8.4

F. Year	2006	2007	2008	2009	2010	2011	2012	2013	2014	Total
Power Loss (MWh)										
Case-A	6,168	6,168	6,168	6,168	6,168	6,168	6,168	6,168	6,168	120,838
Case-B	4,191	4,191	4,191	4,191	4,191	4,191	4,191	4,191	4,191	82,410
Case-C	262	262	262	262	262	262	262	262	261	4,674
Cost of Loss (MMRp)										
Case-A	616.8	616.8	616.8	616.8	616.8	616.8	616.8	616.8	616.8	12,084
Case-B	419.1	419.1	419.1	419.1	419.1	419.1	419.1	419.1	419.1	8,242
Case-C	26.2	26.2	26.2	26.2	26.2	26.2	26.2	26.2	26.2	467
Worth of Cost in the year 1995 (MMRp)										
- Rate -	0.2875	0.2567	0.2292	0.2046	0.1827	0.1631	0.1456	0.1300	0.1161	8.3658
Case-A	177.3	158.3	141.4	126.2	112.7	100.6	89.8	80.2	71.6	4,935.2
Case-B	120.5	107.6	96.1	85.8	76.6	68.4	61.0	54.5	48.7	3,380.1
Case-C	7.5	6.7	6.0	5.4	4.8	4.3	3.8	3.4	3.0	169.7

6-3-4 Plant Site for Power Station

Table 6-11 summarizes construction costs for pipeline system and transmission lines in addition to transmission loss. It is clear from this Table that the construction cost is the least expensive for Sengeti case, followed by the Setiti case and then the Payo Selincah case in this order. However, the overall result including energy loss due to transmission shows that Payo Selincah case is the most economical among the three cases discussed. The following analysis is made based on the case that the power plant will be erected in Payo Selincah.

Table 6-11 Economic Comparison of Plant Site for Power Station

(Unit: Million Rp)

	Case A (Sengeti)	Case B (Setiti)	Case C (Payo Selincah)
Construction Cost			
Pipeline	0.0	1,198.8	2,397.6
Transmission Line	2,445.3	1,683.0	789.0
Sub-total	2,445.3	2,881.8	3,186.6
Power Loss	4,935.2	3,380.1	169.7
Total	7,380.5	6,261.9	3,356.3

All of above comparison are made based on 20 kV transmission voltage which is the standard voltage in Jambi district. Comparison between 20 kV and 70 kV is made and shown in Table 6-12 as a reference. 20 kV transmission line is more economical than 70 kV for a distance of 20 to 30 kilometers.

Table 6-12 Comparison of 20 kV Transmission Line and 70 kV Transmission Line

	If Power Plant is located at Sengeti		If Power Plant is located at Setiti	
	70 kV	20 kV	70 kV	20 kV
Conductor Type	ACSR 300 M C M Al - 150 mm^2	Underground Cable Cu - 400 mm^2	ACSR 300 M C M Al - 150 mm^2	Underground Cable Cu - 400 mm^2
Resistance	6.429 Ω	1.195 Ω	4.689 Ω	0.8716 Ω
Line Length	27.15 km	27.15 km (Water Cable 0.87 km)	19.8 km	19.8 km
Annual Transmission Energy	149,022 MWh	149,022 MWh	131,490 MWh	131,490 MWh
Annual Transmission Losses	2,597 MWh	11,826 MWh	1,475 MWh	6,716 MWh
Initial Capital Investment	11,684 MMRp	2,447 MMRp	9,813 MMRp	1,683 MMRp
Annual Cost of Transmission Losses	260 MMRp	1,183 MMRp	148 MMRp	612 MMRp
Annual Levelized Cost	1,402 MMRp	294 MMRp	1,178 MMRp	202 MMRp
Total	1,662 MMRp	1,477 MMRp	1,326 MMRp	874 MMRp

6-4 Determination of Natural Gas Transmission System and Pretreatment System

6-4-1 Method of Natural Gas Transmission

Natural gas pipeline shall be installed in order to transfer the natural gas produced in Sengeti gas field to Payo Selincah, where the power plant is planned for construction. The pipeline system being planned is expected to transfer a required volume of the natural gas by its inherent pressure without the assistance of compressor. The pipeline route being selected will by-pass highly populated areas as much as possible in order to protect the residential areas' inhabitants and to decrease the pipeline construction cost.

6-4-2 Natural Gas Pretreatment System

(1) Purpose of Pretreatment

In planning the pretreating system for natural gas which will be produced from a production facility and transmitted to designated consumers through pipeline, it is necessary to take the following points into consideration.

(a) Consumer's Requirements for the Gas

There are various usages of natural gas, such as fuel gas for a power plant, city gas, chemical plant feedstock, etc. and the requirements for the gas may also vary depending on each case.

Considering the case in this project that the gas will be utilized as a fuel gas for the dual-fuel engines in a power generation plant in this project, the gas qualities listed below are to be considered.

- Gas Heating Value: heat generated when combusting the unit volume of gas

- Heavy hydrocarbon content in the gas, which will cause knocking of engine and/or a soot generation
- Contents of toxic and corrosive materials contained in gas such as hydrogen-sulfide.

(b) Requirements for Natural Gas Processing Plant

If water and/or carbon-dioxide are contained in a feed gas to a plant in which the feed gas is processed at low temperature, such as LPG (Liquefied Petroleum Gas) or LNG (Liquefied Natural Gas) production plant, icing or carbon-dioxide solidification may occur and cause plugging of pipe and equipment. Therefore, these components shall be removed prior to feeding to the plant.

(c) Requirements for Gas Transmission Pipeline

Followings are commonly considered.

- Dew Point of Natural Gas

  It is necessary to reduce moisture and heavy hydrocarbon content in the gas to prevent their condensation in the pipeline because the associated two phase flow of gas and condensate reduces pipeline transportation efficiency.

  Dehydration is also important to protect the pipeline from internal surface corrosion caused by carbonic acid, condensate dissolving carbon-dioxide gas.

- Gas Hydrate

  Under an elevated pressure condition in the presence of condensed free water, water molecules may react with light hydrocarbon gases to form a solid hydrate, even when operating temperature is higher than a water icing point, i.e., 0°C.

This hydrate formation can be avoided by depressing the moisture content in the gas to a degree that it dose not condense in a gas transmission pipeline system.

Accordingly, from the above discussion, the following systems are to be evaluated as the natural gas pretreating system.

(i) Dehydration process
(water removal)

(ii) Dew point control process
(water & heavy hydrocarbon removal)

(iii) Acid gas removal process
(CO_2 & H_2S removal)

(2) Applicable Processes

(a) Dehydration Process

As dehydration processes: (i) solvent absorption process, using chemicals such as triethylene glycol, diethylene glycol, etc., as solvent; (ii) adsorption process using a molecular sieve or silica-alumina and (iii) self-gas-cooling process by an adiabatic expansion are generally applied.

(b) Dew Point Control Process

(i) Self-gas-cooling process by an adiabatic expansion (an isenthalpic expansion) as same as a dehydration process, (ii) process which cools the gas by an additional external refrigeration and (iii) combination process of (i) and (ii) above, are generally applied as dew point control processes.

During gas processing, there will be same possibility that the temperature of the gas becomes lower than (i) the icing point equal to 0 deg.C or (ii) gas hydrate formation point (this depends on operating conditions). Glycol injection to the upstream of an adiabatic expansion pressure regulating valve or the hot inlet line of a gas cooling heat-exchanger is needed in order to prevent icing of condensed water and a hydrate formation.

(c) Acid Gas Removal Process

For acid gas removal, an absorption process using a circulating solvent is commonly used.

The solvents are classified with the chemical component which is contained in the solvent solution, i.e. (i) chemical solvent with alkanolamine, (ii) chemical solvent with hot potasium, (iii) physical solvent, and (iv) combination solvent of (i) and (iii).

Each solvent has some advantages/disadvantages, therefore a selection shall be made considering the process requirements and the feed/product conditions of gas.

(3) Application to This Project

(a) Necessity of Pretreatment

(i) Consumer's Requirements

According to studies of the chemical compositions of the gas from the existing high pressure separator as the fuel gas for dual-fuel engines for a proposed power generation plant, heavy hydrocarbon exists to some extent but the gas itself is acceptable in its quality.

The carbon-dioxide content is approx. 7 percent and slightly reduces the heating value of the gas but remains in an acceptable range. A reasonable content of the carbon-dioxide can help the engine from knocking; thus, it is not necessary to remove carbon dioxide.

Toxic material such as hydrogen-sulfide is not contained in the gas.

Consequently, no gas treatment is required to meet the consumers' requirement.

(ii) Requirements for Natural Gas Processing Plant

As shown in next Section 6-4, LPG will be recovered from the condensate from the existing high pressure separator, not from a separated gas. Therefore, the LPG recovery plant will not be operated at low temperature, and any pretreating required in low temperature processes will not be necessary.

(iii) Requirements for Gas Pipeline Transmission

The gas from the existing high pressure separator is discharged at dew point condition. If this gas is sent directly into the pipeline, water and/or hydrocarbon condense inside the pipeline upon contact with a cold surface, and may cause the problems which are described in Section 6-4-2, (1)-(c). Therefore, water and heavy hydrocarbon shall be removed prior to sending the gas into the pipeline.

(b) Applied Process

Consequently from the above discussion, the pretreatment system for this project shall be necessary to remove water and control dew point by using adiabatic expansion effect. In addition an acid gas removal process is not required.

(c) Selection of Process

(i) Dehydration Process (water removal)

In a comparison of processes, an absorption process or adsorption process can dehydrate moisture content of the gas to a higher grade than that of self-gas-cooling process by adiabatic expansion; consequently, water and/or hydrocarbon will not condense or gas hydrate will not form even if the temperature surrounding the pipeline becomes low, as is not anticipated in this project area.
However, these processes require a higher investment cost and is not economical as in the case of this project in which the temperature surrounding the pipeline is relatively high.

On the other hand, the self-gas-cooling process by adiabatic expansion can reduce the water level not to the degree of the above processes but to a level acceptable for this project. This process can control the dew point at the same time and its investment cost is relatively lower than the others.

By these reasons, the self-gas-cooling process by adiabatic expansion is selected as the most applicable process for this project.

(ii) Dew Point Control Process (heavy hydrocarbon removal)

Applying the self-gas-cooling process by adiabatic expansion to this project, a dew point of the treated gas can be reduced down to approx. 15 deg.C or less. The minimum expected temperature surrounding the pipeline is 20 deg.C or higher; consequently hydrocarbon will not condense in the pipeline.

Accordingly, supplemental external refrigeration, in order to achieve a higher (lower dew point) grade of gas treatment, is not necessary.

(d) Installation Site

Since this facility is a pretreatment facility to process the gas from GOSP to meet requirements for transmission through pipeline, it is necessary to install this facility between the outlet of GOSP and the inlet of the pipeline. Therefore, the installation place will be near GOSP at Sengeti.

6-5 Determination of LPG Recovery System

6-5-1 LPG Recovery Method

In an LPG Recovery Plant, propane and butane contained in the natural gas are separated and recovered as LPG (Liquefied Petroleum Gas). The recovered LPG is used as fuel for household use and industrial use.

As explained in Chapter 5, the natural gas produced from Sengeti gas reservoirs is separated into gas and condensate at the High Pressure Separator in GOSP. Since the natural gas contains considerable amount of LPG fraction as shown in Table 5-6, both the gas and the condensate from the HP Separator contain a large amount of LPG fraction. Therefore, in order to recover LPG from the natural gas, a process to recover LPG from the HP Separator gas as indicated in Figure 6-8 or a process to recover LPG from the HP Separator condensate as indicated in Figure 6-9 is applicable to this project.

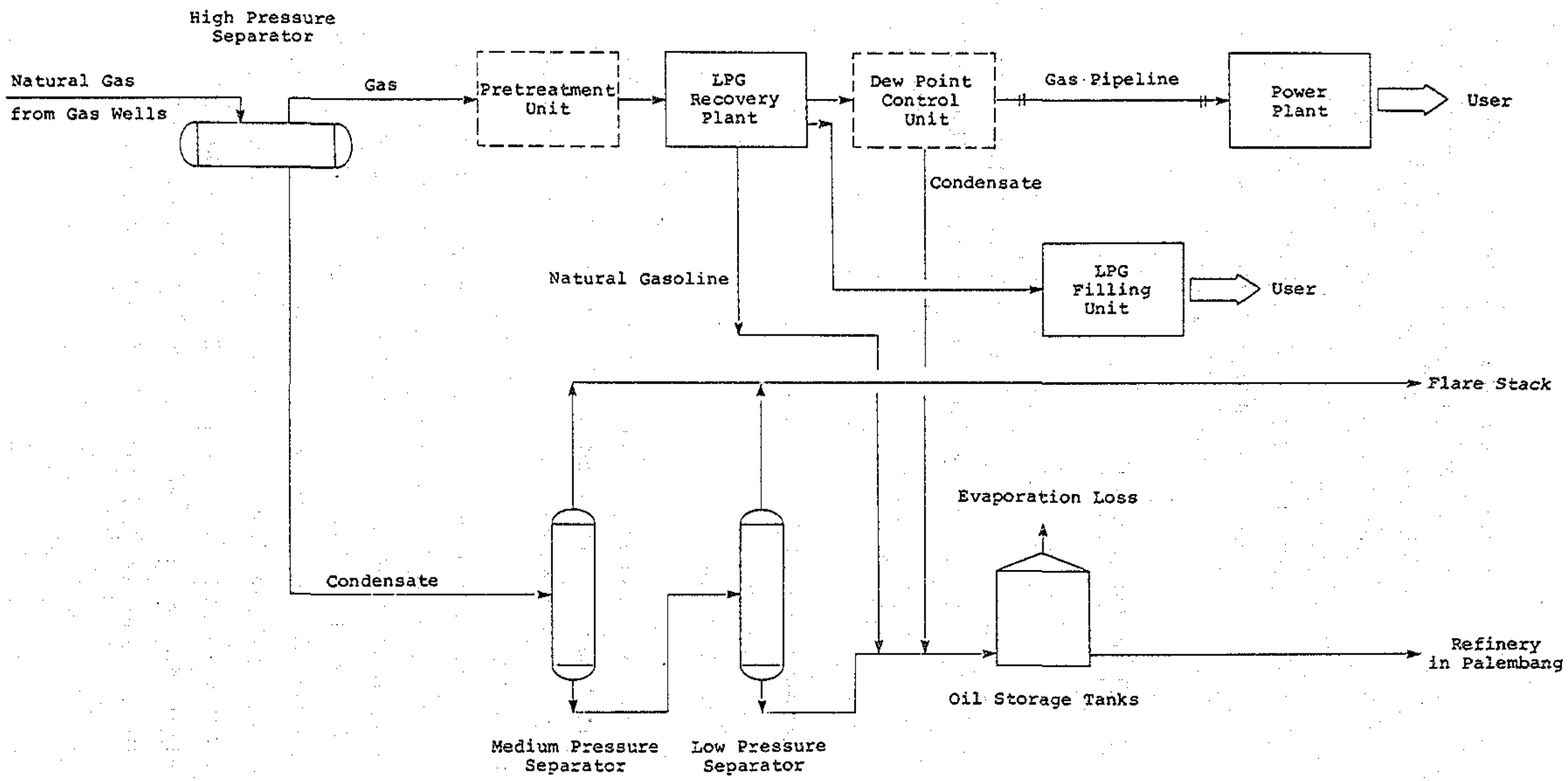

Figure 6-8 Process Configuration of LPG Recovery System from High Pressure Separator Gas

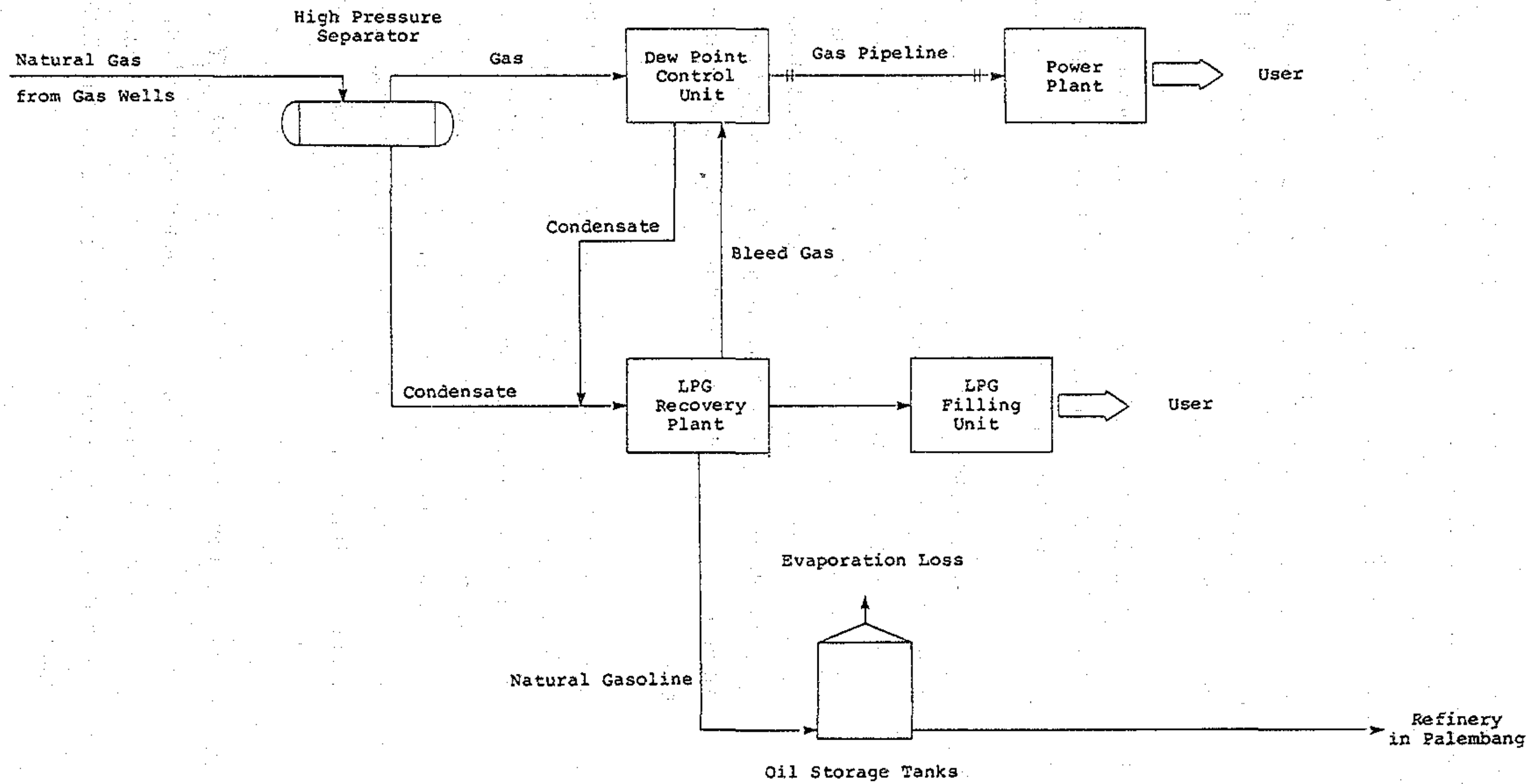

Figure 6-9 Process Configuration of LPG Recovery System from High Pressure Separator Condensate

In view of characteristics of the natural gas, economics, and utilization of resources, adoption of a process to recover LPG from the HP Separator condensate appears appropriate in lieu of a process to recover LPG from the HP Separator gas for the following reasons:

(a) Since LPG contents in the HP Separator condensate are high, it is possible to supply LPG to meet a demand in the region, by using the condensate as feed.

(b) Compositions of the HP Separator gas as shown in Table 5-13 satisfy the required specification of fuel for the proposed power plant, so a facility to adjust compositions of gas is not required.

(c) In the case of LPG recovery from the HP Separator condensate, it is not necessary to install an NGL Recovery Plant which liquefies and separates heavier hydrocarbons contained in natural gas, thus construction cost as well as operation cost are greatly saved, compared with the case of LPG recovery from gas (Refer to Appendix-4). NGL is the abbreviation for Natural Gas Liquid, which is the condensates of heavy hydrocarbon contained in natural gas.

(d) In the case that LPG is not recovered from the HP Separator condensate, the condensate will be processed through Medium Pressure, Low Pressure Separator and sent to the refinery in Palembang via storage tanks. However, most of LPG fraction in condensate will vaporize when stored in storage tanks. Therefore, it is essential to recover LPG from condensate from viewpoints of effective utilization of natural resouces and of air pollution control.

(e) Since the LPG Recovery Plant can be isolated from the supply system of fuel gas for the proposed power plant, the supply volume and compositions of gas to the power plant will not be affected even if there is some trouble in the LPG Recovery Plant.

6-5-2 LPG Production Volume

According to Chapter 4 MARKET STUDY, it is forecast that LPG demand in Jambi district will be 1,540 tons per year and 2,640 tons per year in 1995 and 2000, respectively. The annual operating days will be 335 days, given the maintenance period; therefore, the daily production rate of LPG will be 5 tons to 8 tons per day. Incorporating the demand increase of LPG thereafter, 10 tons per day production will suffice LPG demand in this region for some time. Therefore, the production rate of LPG Recovery plant is tentatively set as 10 tons per day. Design capacity of the plant will be finally determined by optimizing operating conditions in the conceptual design stage. If this amount of LPG is produced, the plant will exceed the regional demand for 7 to 8 years after the completion of the plant. Surplus of production will be sent to Palembang and production adjustment is not contemplated (For details, refer to Chapter 16 *FINANCIAL ANALYSIS*).

6-5-3 Selection of Plant Site

The following items shall be evaluated in selecting a site for the LPG Recovery Plant.

- Supply of raw materials
- Transportation of products
- Acquisition of land
- Construction work
- Effective use of auxiliary facilities
- Soil conditions
- Environmental control and safety precautions for residents

As established during the field survey, Sengeti and Setiti are selected as an installation site for the LPG Recovery Plant and the comparison of the two sites is shown in Table 6-13.

Table 6-13 Comparison of Candidate Site for LPG Recovery Plant

	Sengeti	Setiti
Supply of Raw Materials	◎	△
Transportation of Products		
LPG	o	◎
Light Hydrocarbon	◎	△
Natural Gasoline	◎	◎
Acquisition of Land	◎	◎
Construction Work	◎	◎
Effective Use of Auxiliary Facilities	◎	x
Soil Conditions	o	◎
Environmental Control and Safety	◎	◎

Note:

◎ : Excellent

o : Good

△ : Fair

x : Poor

As indicated in the Table, one of the major deciding factors in selecting one site among the two locations is transportation of raw materials and products. Raw material condensates are supplied from GOSP located at Sengeti and the pretreatment facilities planned to be built near by. Therefore, Sengeti is superior to Setiti from a viewpoint of supply of the raw materials. If the LPG Recovery Plant is constructed at Setiti, it is necessary to install a pipeline to transfer condensate from Sengeti. The existing oil pipeline may be considered for condensate transfer, however, an operating pressure to transfer condensate is high compared with transfer of crude oil, thus the existing pipeline cannot be utilized unless its design pressure exceeds 50 kg/cm^2G. In terms of product transportation of LPG filled in cylinders, Setiti is preferable, where the road for trucks is paved and is located close to Jambi City where majority of LPG are consumed.

When the bridge currently under construction over the Batang Hari river is completed and roads to and from Jambi City are constructed, Sengeti may not be disadvantageous compared to Setiti. It is desirable to feed a by-product light gas to the proposed power plant in view of utilization of natural resources. However, since a fuel gas pipeline installation is planned for the Sengeti side of the Batang Hari river, it is necessary to install an additional pipeline to return the by-product light gas to Sengeti side, if the LPG Recovery Plant is constructed at Setiti. Otherwise, the by-product light gas shall be flared out. The other major factor in selecting a plant site is the utilization of auxiliary facilities such as utility facilities. In each of GOSP and LPG Recovery Plant, the required facilities are a power generating facility, instrument air supply facility, fire fighting facility, flare stack to burn off-gas, waste water treatment facility, control room, office building and warehouse. However, if LPG Recovery Plant is installed adjacent to the existing GOSP, all of those facilities can be shared and the construction cost can be reduced considerably. In view of the above, Sengeti is superior to Setiti as the candidate site of the LPG Recovery Plant.

6-6 Overall System Evaluation

It is necessary to plan and coordinate an overall system as well as pay due consideration to safety measures and environmental protection measures, after integrating individual capacity of facilities as described above, in order to produce electricity and LPG as final products steadily and safely without polluting.

6-6-1 Utilization of Natural Resources

Heavy hydrocarbon condensates are removed for the purpose of adjusting dew point of the gas in the natural gas pretreatment facility and the light gas will be drawn off from the LPG Recovery Plant in order to adjust vapour pressure of LPG product. It is re-commendable to utilize these by-products in the interests of using natural resource and of environmental protection. The methods for these utilizations are described below:

(1) Heavy Hydrocarbons from Natural Gas Pretreating Facility

The heavy hydrocarbons (condensate) contain as much as 60% LPG fraction and are sent to LPG Recovery Plant for LPG recovery.

(2) Light Gas from LPG Recovery Plant

Since the light gas is mainly consisted of C1, C2 and C3 and does not contain heavy components, it will be introduced into the fuel gas pipeline for the proposed power plant. By the inclusion of the light gas into the fuel gas, the dew point of the fuel gas will be raised by about 0.5°C, but this will not pose any problem for pipeline transmission because there is some allowance for setting a dew point at the dew point control facility.

6-6-2 Allowance of Capacity of Facility

Capacity of each facility is established based on the demand forecast of the final products and characteristics of the natural gas.
Since both gas properties and demand forecast are developed based on various assumptions, the capacity defined as above may differ from actual values. Therefore, it will be required to have some allowance for some of facilities to absorb difference of capacity figures. In evaluating allowance of capacity, each facility can be grouped into 3 categories as follows:

- Natural gas production system (wellhead, GOSP)
- Natural gas treatment system (natural gas pretreatment facility, gas pipeline, power plant)
- LPG Recovery system (LPG Recovery plant, LPG filling station)

As explained in Section 6-1, since the governing factor to influence overall capacity is the capacity of the power station, the natural gas treatment system is discussed first.

(1) Natural Gas Treatment System

As indicated in Section 6-2-1, the generating capacity of the power plant is:

Rated output	20 MW
Normal output	17 MW

If the power plant is used to provide a base load, its generating amount is normally 17 MW. Therefore, normal gas processing capacity of each facility which consists the entire natural gas treatment system should be capable of processing fuel gas corresponding to generating capacity of 17 MW. If the maximum gas processing capacity corresponding to 20 MW generating capacity is used as the basis of plant design, the operation should possess the flexibility to cope with changes of boundary conditions.

Therefore, gas processing capacity of each facility is set as follows:

Normal capacity:

Capacity to process the fuel gas corresponding to generating output of 17 MW

Maximum capacity:

Same as above, but corresponding to 20 MW

(2) Natural Gas Production System

The role of the natural gas production system is to supply the gas of a uniform quality and quantity to the natural gas treatment system. However, as explained in Chapter 5, the compositions of the natural gas produced from the reservoirs will differ from the set value and will be changed during operation by change of the producing wells. Therefore, in order to cope with changes of gas compositions, it is necessary to plan to have some flexibility of the operating pressure for High Pressure Separator.

In other words, if heavy cuts of hydrocarbon increase in natural gas, operating pressure is raised to condense heavy cuts and if heavy hydrocarbon decreases on the contrary, the operating pressure is reduced, so that High Pressure Separator gas can be adjusted to a uniform quality. Since the existing facility of natural gas production system is utilized, actual operating pressure range will be 20 to 40 kg/cm^2G (the designated operating pressure is 25 kg/cm^2G). Thus, the operating pressures for natural gas pretreatment facility and gas pipeline system may increase beyond the designated operating pressure and some countermeasures are required (refer to Section 6-5-2).

By adjusting the operating pressure of High Pressure Separator, gas volume from HP Separator is changed. In order to keep fuel gas volume constant, gas volume of feed gas also needs to be adjusted. However, judging from compositions of the gas reservoirs of 10 layers as discussed in Chapter 5, it is anticipated that amount of adjustment for natural gas supply will be minor as there is not much difference in the compositions of each layer. Therefore, the capacity of natural gas production system and the capacity of the natural gas treatment system are defined as follows,

Normal capacity:

Capacity to process fuel gas corresponding to the generation output of 17 MW

Maximum capacity:

Same as above, but corresponding to 20 MW

(3) LPG Recovery System

As explained in (2) above, variations of composition of feed natural gas is adjusted by changing the operating pressure of High Pressure Separator and by changing feed volume of the natural gas; and the consequent result is observed as the variation of composition and flow rate of condensate from High Pressure Separator to LPG Recovery Plant. Therefore, it is necessary to have a wide range of operations in LPG Recovery Plant in which High Pressure Separator condensate is fed as a main feed. The operating range will be determined in the conceptual design stage after the specifications of each piece of equipment are known.

6-6-3 Maximum Operating Pressure of Facility

Due to changes of operation of High Pressure Separator to cope with the changes of feed natural gas, the operating pressures of the natural gas pretreatment facility and of the gas pipeline change accordingly. However, since the light gas from deethanizer of LPG Recovery Plant will be injected into gas pipeline by the inherent self-pressure, it is mandatory to keep the operating pressures of the natural gas pretreating facility and of the pipeline below the operating pressure of the deethanizer. Therefore, the maximum operating pressure for these two facilities will be determined in relation to the conceptual design of LPG Recovery Plant.

A pressure regulating valve will be installed at the inlet of the power plant, and the operating pressure of fuel gas supply system in the power plant can be set independently from the gas processing facility.

6-6-4 Spare Equipment

Spare units of the rotating machines which run continuously will be provided so the entire system will not be affected by any unexpected operation trouble.

6-6-5 Safety Measures

Natural gas and LPG handled in this project are combustible and explosive materials. Therefore, it is mandatory to secure the safety of operating personnel and residents nearby; accordingly, the following safety measures shall be provided.

(1) Safety Measures for Natural Gas Production and Treatment System

Since high pressure natural gas is processed through these systems, operating pressures are planned so they do not exceed the design pressures of component equipment. Blow-down pipe with a pressure regulating valve will be provided for this purpose and if the operating pressure exceeds a set pressure, gas will be released through it and flared at off-gas burning facility.

(2) Safety Measure for Pipeline

(a) The gas pipeline is planned to be installed underground away from areas where population density is high. In order to protect the pipeline, a protective pipe will be provided for the section across roads, and the part crossing the Batang Hari river will be buried in ditches dug in the river bed floor.

(b) Adequate anti-corrosion measures will be taken. Block valves will be installed to divide the pipeline into several sections so that any portion showing defective signs will be isolated and the gas remaining in that portion of pipeline will be vented within a short period of time.

(3) Safety Measure for LPG Recovery System

(a) Safety valves will be provided at distillation towers so that the operating pressure will not exceed the design pressure, and in the case of an emergency, the gas contained in the system will be discharged to the off gas burning facility.

(b) During the filling operation of LPG into cylinders, gas leakage from head caps of cylinders will be checked by a leakage detector.

(4) Safety Measure for Power Plant

(a) Two safety shut-off valves in series and a spark arrester will be installed in order to isolate the fuel gas supply system and to prevent the flame in the engine from igniting the gas in the pipeline.

(b) Supply pressure of fuel gas to the power plant which requires low pressure gas will be reduced to the designated pressure by a pressure regulating valve installed at the inlet of the power plant. If the pressure of fuel gas deviates from the set value, operation mode will be automatically changed to diesel operation mode (the operation which needs only diesel oil) to protect the engine.

(c) Ventilation mechanism will be provided in places where gas pipeline and auxiliary facilities are installed.

(5) Fire Fighting Facility

Fire fighting facilities will be installed at GOSP, natural gas pretreating facility, LPG Recovery Plant and power plant as safety measures against fire.

(6) Communication Facility

Wireless communication facility will be installed for close co-ordination of operations between Sengeti area and Payo Selincah area, thus, in case of an emergency, operators of both locations are able to communicate with each other immediately.

6-6-6 Pollution Control

The followings are possible pollutants to contaminate environment in implementing this project.

1) Waste Gases

- Nitrogen oxides
- Sulfur oxides
- Dust

2) Waste Water

- Oily water
- Sanitary water from human activities
- Waste oil and sludge

3) Noise

4) Vibration

Sources for above pollutants are:

1) Natural gas processing and LPG handling facilities at Sengeti

2) Power plant at Payo Selincah

Necessary measures for each area are as follows:

(1) Sengeti Area

A natural gas pretreating facility and LPG Recovery Plant are to be constructed adjacent to GOSP which exists in Sengeti. Pollutants discharged from these facilities are oily water and sanitary water from human activities.

(a) Oily Water

Water contained in natural gas produced from gas reservoirs will be condensed in these facilities and be discharged to the Batang Hari river. However, since the water contains oil, it will be treated in the existing oil water separation plant and oil will be separated before the water is discharged to the river.

(b) Sanitary Water from Human Activities

Sanitary water of employees working for these facilities will be treated in a septic tank before being discharged to the river.

Besides waste water, pollutants are discharged from machines operated by diesel engines.

Since these pieces of machines are small in capacity, installed away from residential area and not used frequently, special measures for exhaust gas, noise or vibration will not be necessary.

(2) Payo Selincah Area

Duel-fuel engines will be used for power generation at the power plant planned to be constructed in Payo Selincah area, which consume both natural gas and fuel oil and there will be some degree of environmental pollution of exhaust gas, waste water, noise and vibration.

(a) Exhaust Gas

Air will be polluted by exhaust gas from the engine which contains nitrogen oxides, sulfur oxides and dust. However, there is no plant around the candidate site of the power plant except the existing Payo Selincah power station, and no heavy traffic is observed or anticipated in this area. Therefore, it will not be necessary to consider any special measure for the above. There will be detailed discussion for the amount of exhaust gas in the conceptual design.

(b) Waste Water

Since waste water from the power plant will contain fuel oil and lubricating oil, the oil in water will be separated in a waste water pit before being discharged. Recovered oil, deteriorated fuel oil, and lube oil shall be burned in an incinerator.

(c) Noise

Noise from engines will not be nuisance if the engines are installed inside a building located away from neighbouring residents.

(d) Vibration

Vibration from engine will be absorbed in a specially designed foundation.

6-7 Optimum Project Scheme

The optimum project scheme determined by the process described above is summarized below:

(1) Power Plant

Generating capacity: Rated: 20 MW
Normal: 17 MW
Generation system: Dual-fuel engine generator
Site: Payo Selincah area

(2) Natural Gas Pretreating Facility

Gas processing volume:
Maximum - fuel gas corresponding to 20 MW generation
Normal - Same as above, but corresponding to 17 MW
Process: Dew point control by adiabatic expansion self cooling method
Site: Sengeti area

(3) LPG Recovery Plant

LPG production volume:
about 10 tons per day (Details will be decided in the conceptual design)
Process: LPG recovery by distillation of High Pressure Separator condensate from GOSP
Plant site: Sengeti area

Figure 6-10 shows locations of power plant, natural gas pretreatment facility and LPG Recovery Plant, established in the optimum project scheme.

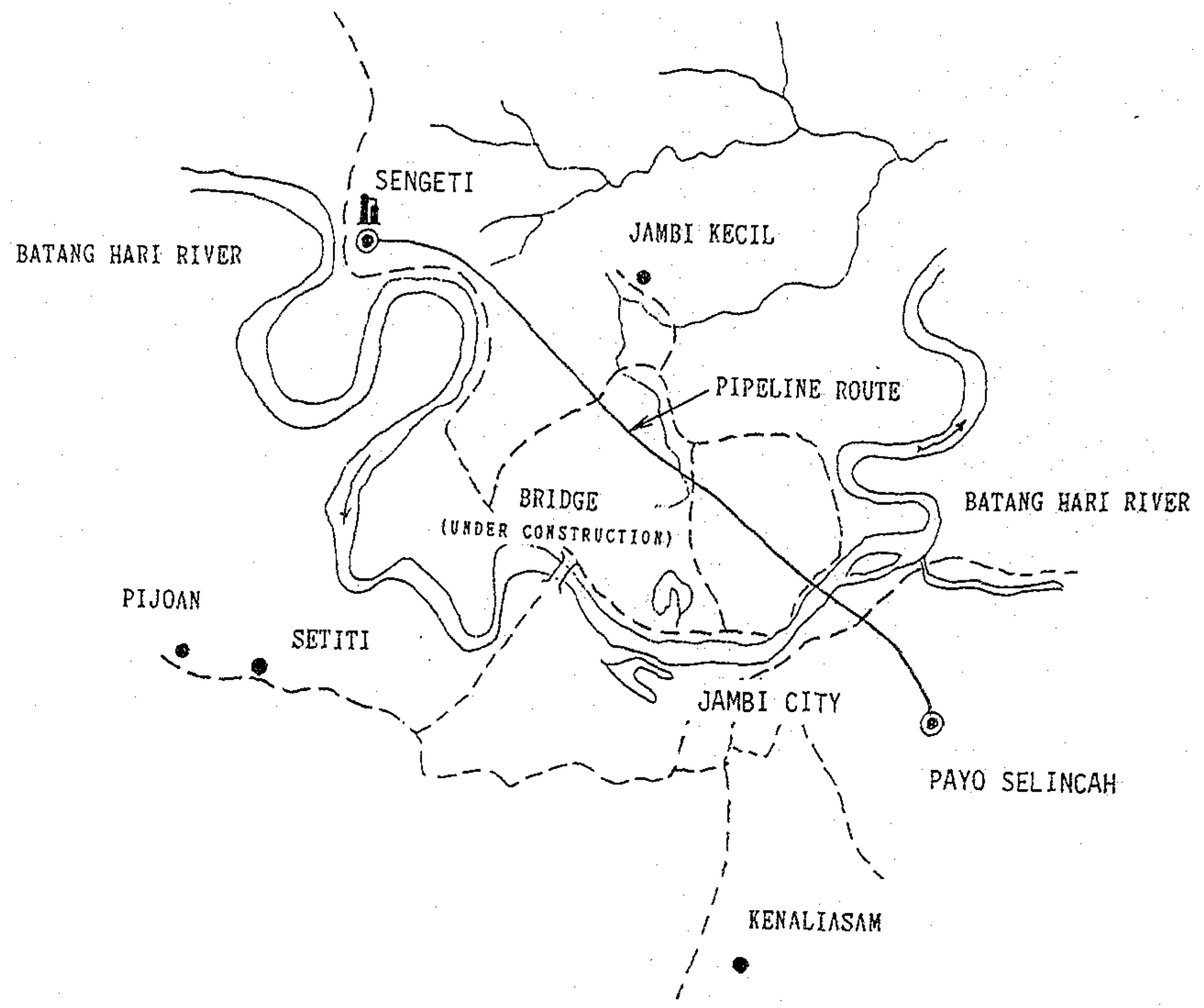

Figure 6-10 Location of Plants

7. CONCEPTUAL DESIGN OF NATURAL GAS PRETREATMENT FACILITY

This chapter describes a conceptual design of the natural gas pretreating facility which processes natural gas produced in Sengeti gas field and produces the pipeline gas to be transmitted to the power plant as its fuel gas.

7-1 Design Condition

The site survey for the Jambi natural gas utilization project was conducted in February, 1988. In consideration of the survey results, design conditions mentioned below have been taken as the design basis for planning this pretreating facility.

It is thereby necessary to partially repair and/or modify the existing GOSP (gas oil separation plant) facilities.

7-1-1 Natural Gas Supply Source

Regarding a gas production from the Jambi gas field, it is planned that the existing GOSP facilities including flow lines and production manifolds will be utilized by carrying out partial repairing work and/or minor modification. Assuming that the utilization of the existing GOSP facilities be available, it has been decided as a design basis that the natural gas supply source to the natural gas pretreating facility will be the gas separated at the existing high pressure separator.

7-1-2 Quality of Feed Natural Gas

Separated natural gas in the existing GOSP will be supplied to the natural gas pretreating facility, where its water and hydrocarbon dew point is controlled, and then finally transmitted to the power plant through the gas pipeline.

A quality of the feed natural gas, which is shown in Table 7-1, already meets the required specification of the fuel gas for the power plant dual-fuel engines.

In this viewpoint, the feed natural gas needs no further treatment for the power plant fuel gas. However, its dew point shall be controlled in this pretreating facility to prevent the troubles during pipeline transmission.

In controlling its dew point, a portion of heavy hydrocarbon is condensed and removed from the gas phase, therefore the quality of treated gas will become better than that of the feed gas shown in Table 7-1. The treated gas will be transmitted to the power plant.

Table 7-1 Quality of Feed Natural Gas

Components	Mole %
C_1	57.1
C_2	15.4
C_3	13.5
iC_4	2.5
nC_4	3.3
iC_5	0.8
nC_5	0.5
C_6 & heavier	trace
CO_2	6.7
H_2O (saturated)	0.2
M.W.	27.05

Conditions	
Pressure	25 kg/cm^2G
Temperature	26.7°C

7-1-3 Utilization of Existing Facilities

As mentioned in the section 7-1-1, the existing GOSP facilities are to be functioned to supply feed natural gas for this plant. The natural gas pretreating facility requires a flare blow-down system and an oily water treating system. The existing GOSP has those systems which are utilized for the natural gas pretreating facility from an economical point of view.

7-1-4 Estimated Minimum Gas Temperature During Pipeline Transmission

Pipeline temperature is an important factor for gas pipeline transmission. Moisture and hydrocarbon condensates which are formed by a temperature decrease will not only lower the gas transmission efficiency but also cause a corrosion problem. Especially moisture condensate may result in a severe corrosion in the presence of CO_2 gas contained in the pipeline gas, therefore, the dew point of the gas shall be controlled in the gas pretreating facility before the gas is sent to the gas pipeline.

Since the pipeline is planned to be installed underground, the temperature of transmission gas is closely related to an underground temperature and finally becomes to the same temperature as the underground soil. It is therefore assumed that the minimum temperature of transmission gas will become almost same to the minimum underground soil temperature.

By controlling the gas dew point below the minimum transmission gas temperature in this plant, the troubles during pipeline transmission can be avoided.

As the gas pipeline will be installed underground 1.0 to 1.2 meters below grade, the minimum underground soil temperature would be at a range between an average minimum ambient temperature (20°C) and an average ambient temperature (26.5°C).

This facility is so designed that the treated gas will not have condensate even at the condition that the transmission gas temperature may decrease to the average minimum ambient temperature (20°C) to have allowance in operation. Thereby, as the design basis of this facility, the minimum transmission gas temperature is set at 20°C.

7-1-5 Application of Skid Mounted Facilities

An application of skid mounted installation is to be considered for the proposed natural gas pretreating facility taking into account the following advantages:

(1) Minimizing of a site construction work
(2) High level quality control by shop fabrication
(3) Easy mobilization to other places for similar purpose
(4) Minimizing of a total plant cost

7-2 Natural Gas Pretreatment Facility

This facility is to control the dew point of natural gas applying self-cooling effect by an isenthalpic expansion (Joule-Thomson expansion). In this facility, natural gas is lowered its dew point to approximately 15°C at the outlet of the facility and is sent to the gas pipeline as a power plant fuel gas. Followings are the brief description of this facility.

7-2-1 Design Consideration

(1) Design Gas Flow Rate

Design gas flow rate shall be decided based on the fuel gas demand for the power plant dual-fuel engines (refer to Chapter 10) and utility gas demand for the LPG recovery plant (refer to Chapter 8).

The fuel gas demand shall be calculated based on the maximum power plant output capacity (5 MW x 4 = 20 MW) as follows:

$$\frac{20(\text{MW}) \times 1{,}000 \times 860(\text{kcal/kWh}) \times 0.9}{0.36 \times 12{,}580\ (\text{kcal/Nm}^3) \times 0.96} = 3{,}561\ \text{Nm}^3\ (3.2\ \text{MMSCFD})$$

Fuel gas ratio (fuel gas/fuel gas + oil)	:	0.9
Engine efficiency	:	36%
Generator efficiency	:	96%

And the required utility gas quantity will be 54 Nm^3/h (refer to Chapter 8), therefore, the total required fuel gas will be 3,615 Nm^3/h. The total required fuel gas of 3,615 Nm^3/h will be obtained by adding both the dew point controlled gas from this facility and the off gas from the LPG recovery plant (refer to Chapter 8). Since the dew point of the off gas from the LPG plant is so low that it does not need any treatment.

190 Nm^3/h of off gas will be supplied from the LPG recovery plant, therefore, the amount of gas required to be dew point controlled and supplied from this gas pretreating facility will be:

$$3{,}615 - 190 = 3{,}425\ \text{Nm}^3/\text{h}\ (3.1\ \text{MMSCFD})$$

In order to produce 3,425 Nm^3/h of dew point controlled gas, 3,438 Nm^3/h of feed gas will be required. Figure 7-1 shows the material balance mentioned above.

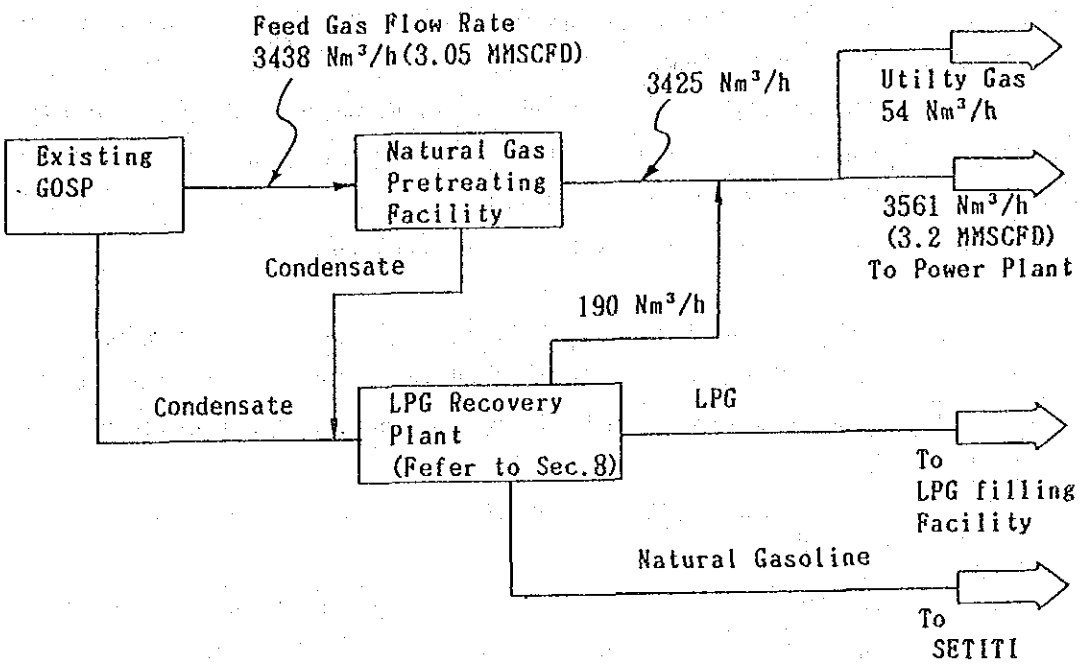

Figure 7-1 Material Balance

(2) Outlet Pressure of Treated Gas

The treated gas at this facility is transmitted to the power plant through a pipeline. The outlet pressure (the pipeline inlet pressure) is decided in consideration of the required pressure of the power plant, which is equal to the pipeline outlet pressure, the flow rate of treated gas, pipeline size and distance.

6 inch pipeline is to be adopted for this project (refer to Chapter 9). In this case, the minimum required pipeline inlet pressure, at the condition that a pipeline outlet pressure of 10 kg/cm^2G, will be 13 kg/cm^2G in order to transmit 3.2 MMSCFD of the treated gas.

On the other hand, the total required differential pressure for applying an isenthalpic expansion effect for the dew point control as well as processing the gas in this facility will be approximately 9 kg/cm^2. Therefore, in case the gas inlet pressure to this facility is 25 kg/cm^2G, the outlet pressure will be:

$$25\ kg/cm^2G - 9\ kg/cm^2 = 16\ kg/cm^2G$$

In this case 3.2 MMSCFD of gas can be transmitted, since the pipeline inlet pressure is higher than 13 kg/cm^2G.

Taking consideration the above, it is decided that the treated gas outlet pressure is 16 kg/cm^2G at which a maximum gas transmission capacity will be approximately 5.2 MMSCFD. For details, please refer to Chapter 9, CONCEPTUAL DESIGN OF NATURAL GAS PIPELINE.

(3) Dew Point of Treated Gas

This facility is designed to control the dew point of the treated gas to the level of 5°C lower than the minimum pipeline transmission gas temperature. By doing so, the gas is always kept at the super heated condition, thus the gas of super heated condition will not form condensation and accordingly troubles during gas transmission are not anticipated. It is thereby decided that the dew point temperature of the treated gas at the outlet pressure (16 kg/cm^{2}G) of this facility will be:

20°C (Minimum transmission gas temperature) - 5°C = 15°C
(Refer to section 7-1-5.)

For reference, a relation between treated gas pressure and dew point temperature is shown on the following Figure 7-2.
In the mean time, at a down stream of the pipeline where an operating pressure lower than 16 kg/cm^{2}G be anticipated due to the pipeline pressure drop, the gas dew point temperature becomes below 15°C and thus gas enters into the condition far away from its condensing point.

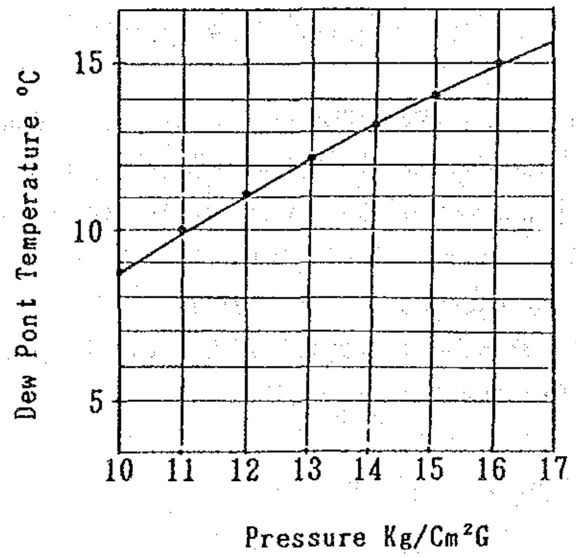

Figure 7-2 Gas Pressure and Dew Point Temperature

(4) Hydrate Inhibition

In this pretreating facility, the gas will be cooled down to approximately 16°C using an isenthalpic expansion to produce the treated gas having a dew point temperature of 15°C at 16 kg/ cm^2G (pipeline inlet pressure). 16°C will be the lowest temperature achieved by this facility when the operating pressure is 17 kg/cm^2G. A hydrate formation temperature at 17 kg/cm^2G is estimated to be 11°C as contemplated from the following Figure 7-3, therefore, no hydrate formation will be anticipated in the normal operation because the lowest temperature of 16°C is high enough comparing to the hydrate formation temperature (11°C).

However, a methanol injection system shall be provided in this facility to prevent the hydrate formation in case of an abnormal operation.

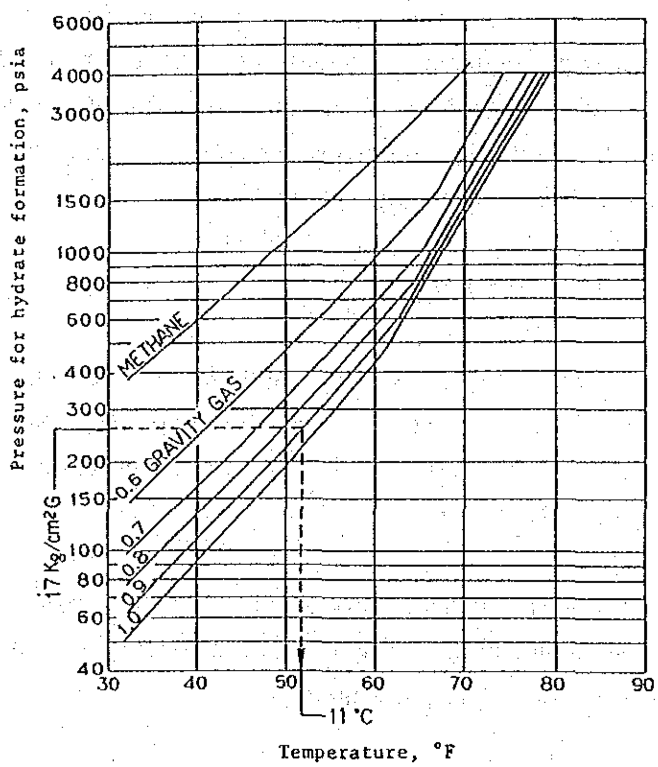

Figure 7-3 Pressure-temperature Curves for Predicting Hydrate Formation

(5) Required Treated Gas Quality

As mentioned in section 7-1-3, the feed gas to this facility inherently meets the required fuel gas specification for the power plant dual-fuel engines.

Accordingly, only dew point control for preventing the troubles during a pipeline gas transmission is the requirement to this facility.
The treated gas of this facility will have better quality for the power plant dual-fuel engines than that of the feed gas because a portion of heavy hydrocarbons contained in the feed gas is removed by a dew point control in this facility.

(6) Design Pressure

In the fuel gas supply system consisting of GOSP (existing), the gas pretreating facility and a pipeline, an operational flexibility shall be required to meet the variation of feed natural gas quality. Therefore, a design pressure of 38 kg/cm^2G which is higher than the specified operating pressure is taken for the entire fuel gas supply system.

(7) Material Selection

Stainless steel is to be used as a tube side material of the heat exchanger which cools the feed gas, because the tube side would be in a corrosive condition due to presence of moisture condensate and CO_2 gas.

For all the other parts, other than the tube side of the heat exchanger, carbon steel will be basically used with an appropriate corrosion allowance.

7-2-2 Treated Gas Quality

The gas from the existing GOSP will be introduced to the pipeline after controlling its dew point in this facility. Table 7-2 shows the treated gas quality.

Table 7-2 Treated Gas Quality

Components	Mole %
C_1	57.6
C_2	15.4
C_3	13.4
iC_4	2.4
nC_4	3.3
iC_5	0.8
nC_5	0.4
C_6 & heavier	trace
CO_2	6.7
H_2O (super heated)	0.1

Conditions

Molecular weight	26.7
Flow rate	3,371 Nm^3/h
Pressure	16 kg/m^2G
Temperature	21.2°C
Dew point (at 16 kg/cm^2G)	15°C

7-2-3 Process Description

This section explains the process flow of natural gas pretreating facility shown on the flow sheet in the following section 7-2-4.

Natural gas separated from the existing high pressure separator in the GOSP is supplied to this facility at approximately 27°C and 25 kg/cm^2G. The gas which is saturated with water and hydrocarbon at 27°C and 25 kg/cm^2G is then sent to the tube side of E-1 (Feed/Product Gas Heat Exchanger) in which it is cooled to approximately 24°C by heat-exchanging with a cold gas and a portion of moisture and heavy hydrocarbon in the gas will be condensed.
The gas is then expanded through a pressure regulating value with reducing its pressure to 17 kg/cm^2G and as a result, cooled to approximately 16°C by an isenthalpic expansion effect, and finally sent to D-2 (Low Temperature Separator).

At D-2, it is separated to gas, hydrocarbon condensate and water. Separated hydrocarbon condensate is pumped to the LPG recovery plant (refer to Chapter 8) newly installed and separated free water is sent to the existing oily water system for treating.

On the other hand, separated gas from D-2 is sent to D-5 (Product Gas Mist Separator) in which the mist be removed, then, introduced to E-1. At E-1, the separated gas is heated from 16°C to 21.2°C by heat exchanging with the feed gas and sent to the pipeline at a super-heated condition.

The dew point of the super heated gas will be 16°C at 17 kg/cm^2G and will be lower than 15°C at the pipeline inlet pressure (16 kg/cm^2G).

In case that the feed gas temperature is higher than 27°C, following operation procedures shall be taken to maintain 16°C at D-2.

(1) The operating pressure of the existing high pressure separator in the GOSP shall be increased up to a certain pressure more than 25 kg/cm^2G.

By doing so, the dew point of 16°C can be maintained because a larger cooling capacity can be taken due to a larger differential pressure obtained at the pressure regulating valve.