

b) Test results of rock samples

Sample No.	1	2	3	4	5
Test item	(blue clay)	(tuff)	(blue clay)	(blue clay)	(tuff)
Unconfined compressive strength (kg/cm <sup>2</sup> )	11.6	8.9	7.5	30.1	4.8
Density (Natural) (g/cm <sup>3</sup> )	2.154	1.824	1.995	2.116	1.910
Natural moisture content ratio (%)	15.9	24.9	21.6	7.0	26.6
P wave (m/sec)	1961	1253	1725	1364	1266
S wave (m/sec)	984	—	890	682	617
Collapse value in water (rank)	D	D	D	D	D

These test results show that all the samples not only belong to soft rocks but feature marked deterioration. Meanwhile, the wet and dry repeated test, scheduled to be conducted, was abandoned because all the samples turned to be muddy (D rank) during the collapse value test in water.

(2) Conceptual plan of coal mining method

1) Mining parameters

In order to make preliminary cost estimation of coal mining, certain assumptions on mining parameters have been made.

i) Production Requirements

- a) Coal production : 3 million tons/year
- b) Amount of actual minable coal and waste spoil, considering loss factors.

	Coal reserves (x10 <sup>6</sup> ) t	Overburden removal (x10 <sup>6</sup> ) m <sup>3</sup>	Stripping ratio (m <sup>3</sup> :t)
In-Situ	123.0	258.7	(2.10:1)
Actual figure	98.4	277.9	2.82:1
(assuming weathering loss, geological loss, mining loss, 5%, 10%, 5% respectively)			
Mineable coal	98.4 x 10 <sup>6</sup> t		
Coal production	3 x 10 <sup>6</sup> t ÷ 1.28 (sp.gr.) = 2.344 x 10 <sup>6</sup> bank m <sup>3</sup> /year		
Overburden removal	3 x 2.82 = 8.46 x 10 <sup>6</sup> m <sup>3</sup> /year		
Total materials	2.344 + 8.46 = 10.8 x 10 <sup>6</sup> m <sup>3</sup> /year		

c) Coal demand : continuous-constant demand through the year, excepting around June (35 days) for periodical maintenance of coal gasification plants.

d) Mine life : 33 years  
98.4 ÷ 3 = 33

e) Operating days : 302 days/year  
Scheduled working hours are calculated on a 7-day/week, 3-shift/day basis. The 3-shift operation is an international mining working custom.

Excluded are holidays every week numbering 52 days, 11 national holidays. Moslem holiday (only one shift per each Friday) is included in 302 days.

f) Climatic condition

Based on statistics covering a 30-year period, the average annual precipitation in the area is estimated to be 3,147 mm, with an average of 162 rainy days per year. According to mining operation records of the Bt. Asam mine, production (including output of overburden removal) decreases to 53% in rainy season.

## 2) Selection of mining methods

### i) General

Since the Block-A area has fewer overburden removal ( $2.82 \text{ m}^3/\text{t}$ ), slightly inclined seams ( $10^\circ-15^\circ$ ) and multi-seams deposit, bench cut surface mining is the best adaptable mining method.

The same mining method is better to employ in coal mining and overburden removal due to following reasons.

- a) A mining method featuring interchangeability between the rainy and dry seasons is required.
- b) With face floor kept horizontal from the aspects of maintenance and efficiency, a bench irregularly involves both the coal mining and overburden removal works.
- c) Judging from the Bt. Asam mine's experience, there is little difference in digability between coal mining and overburden removal.

In general, the selection of the mining method depends on the local conditions such as geological conditions, mining parameters, technical reliability and financial conditions. However above local conditions are not yet fully elucidated at this time. Therefore, conceptual plan of two different type of mining method, continuous mining method and noncontinuous mining method are examined below to select an optimal, efficient one to the Banko area.

### ii) Outline of continuous mining method

Bucket wheel excavators ... Digging & loading

Belt conveyors ..... Transportation

The B.W.E. is regarded as the most applicable because digability of the objects of mining (particularly overburden) suggests that the need for blasting will be limited within an auxiliary means and that B.W.E.-based cutting will fully be feasible. To transport coal and overburden, the belt conveyor system will be employed as continuous transport equipment common to

faces and overland.

- iii) Outline of non-continuous mining method
  - Rope shovels ..... Digging & loading
  - Rear dump trucks ..... Face transportation
  - Belt conveyors ..... Overland coal and spoil transportation

In regard to cutting machines, large-size hydraulic shovels are disqualified due to their insufficient achievement in operation and rope shovels, verified by rich achievements, is selected as the object to examinations. Because a reasonable face transportation system is vehicle-type one that can match shovel cycle and capacity, well demonstrated rear dump trucks are employed, though belt conveyors are employed in the fixed routes from the outside of faces to coal stock yard and spoil dump area.

3) Conceptual plan of continuous mining method

- Bucket wheel excavators system -

i) Mining system

Fig. 7-3-3 illustrates the proposed mining system and Fig. 7-3-4 shows proposed arrangement of transportation route.

a) Arrangement of faces

The whole area subject to mining is divided along the strike side into five blocks, each measuring some 1,600m in strike side length. Mining is initiated first in the block containing the most stable coal seams and involving the least amount of overburden. Within a block installed are five tiers of 20m-high benches. Five faces are constantly arranged, and advancing toward the same direction in principle, starting from the top bench and followed by lower ones in an one-after-another manner. Four faces are in operation and the remaining one is reserved, while the five faces advance at the same pitch with regulated

Fig 7-3-3 Illustration of Mining System

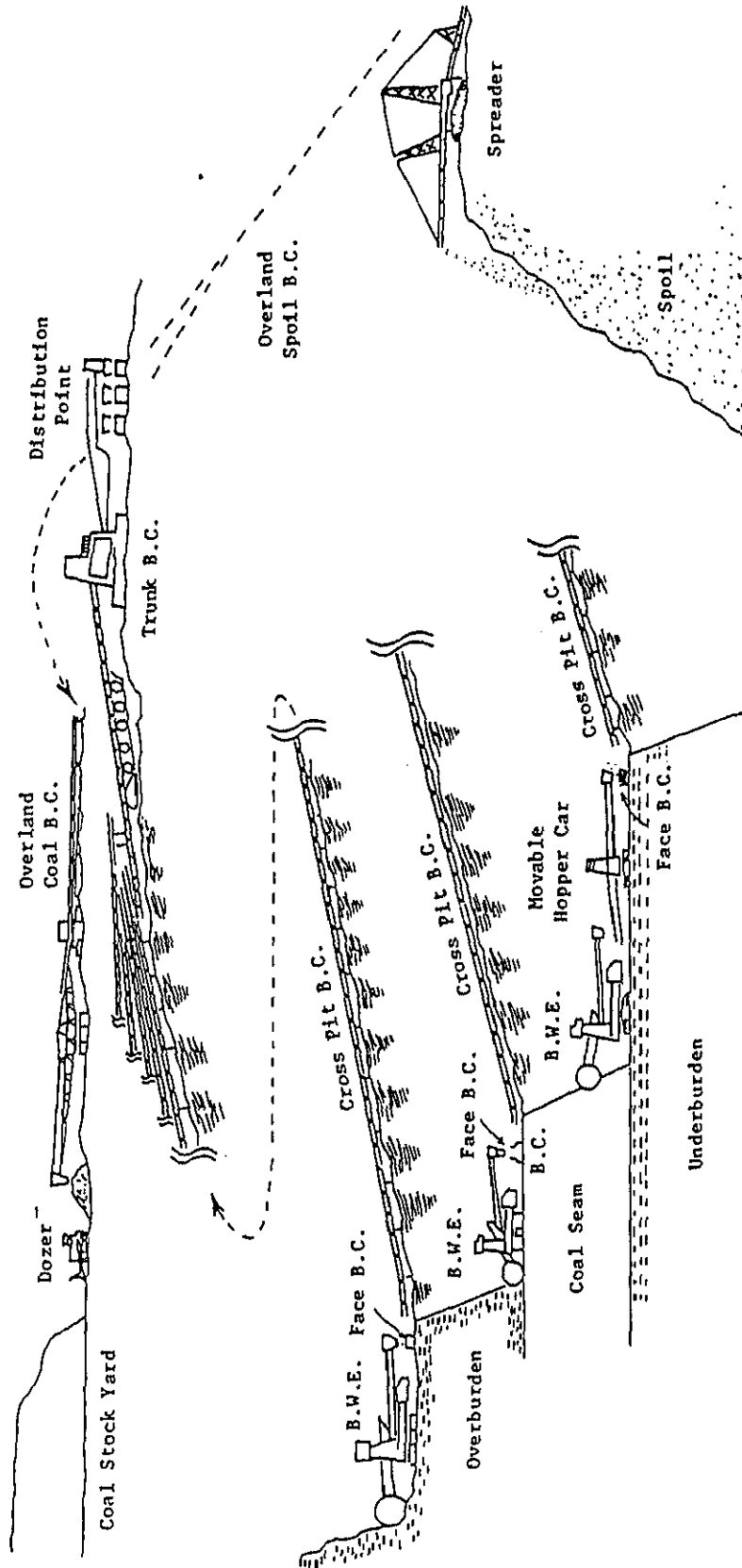
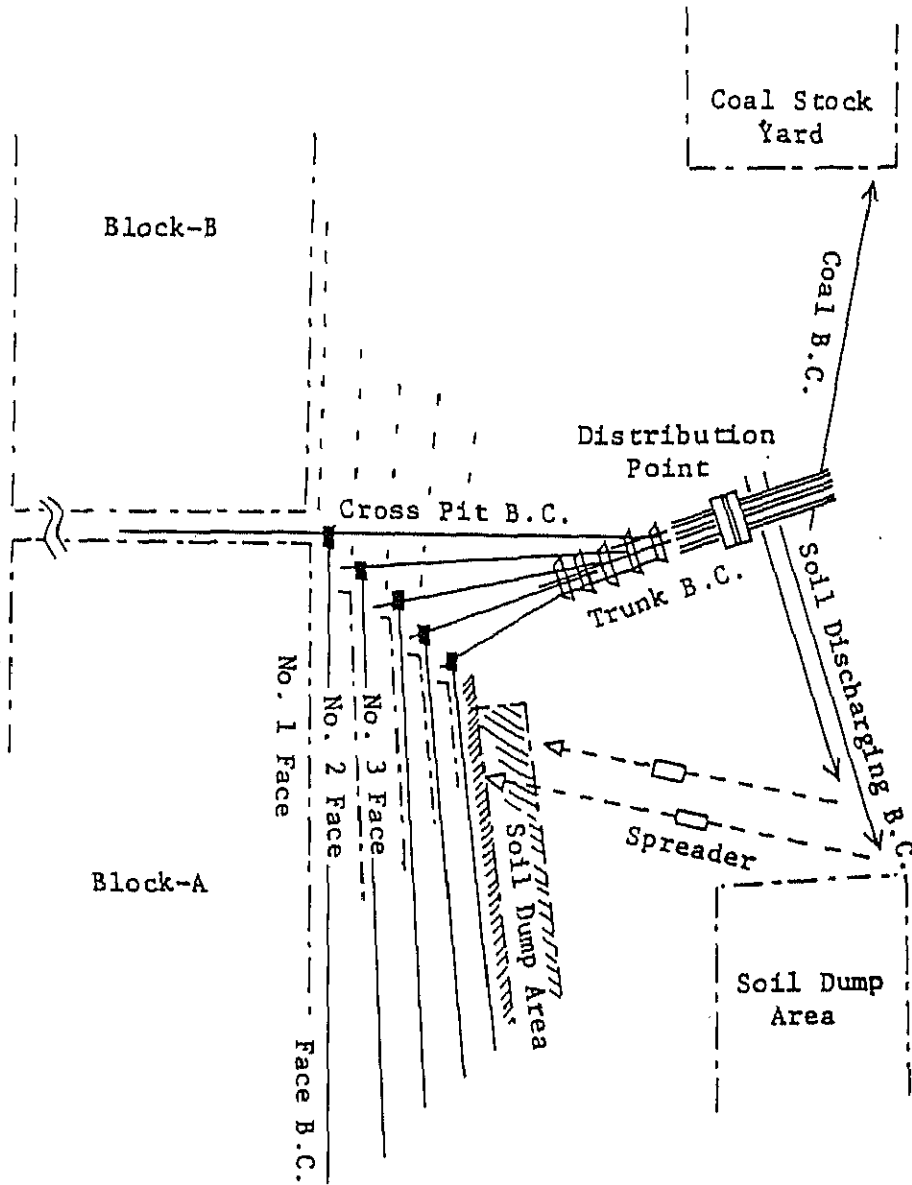


Fig. 7-3-4

General Arrangement of Transportation Routes



mining. Face shifting is due in every bench upon the completion of the first block to the next one.

b) Mining operations in faces

- b-1) Mining: Employing bucket wheel excavators, mining and loading of overburden and coal are conducted.
- b-2) Complementary mining: With scrapers, shovels, dump trucks and dozers employed, new faces, transport routes, equipment shifting routes and others are prepared. Also conducted are cleaning of face floor and maintenance of road surface.
- b-3) Hardrock arrangement: Hard rocks, if any, are bored with a drill rig, then weakly blasted up.

c) Transportation

- c-1) Face: Belt conveyor system with shiftable conveyors and cross pit conveyors.
- c-2) Distributing of overburden material and coal: Trunk belt conveyors system.
- c-3) Overburden discharge: Overland belt conveyors with shiftable conveyors, movable tripper cars and movable spreaders.
- c-4) Coal transportation: Overland belt conveyors.

ii) Estimated actual driving hours of mining equipment

a) Factors taken into consideration

- a-1) Coal demand: needs for continuous loading to gasification plant except periodical inspections and maintenance in the dry season.
- a-2) Labour holidays: weekly holidays, national holidays and Moslem holidays.
- a-3) Spontaneous combustion of coal while stored: limitation of stock-pile interval
- a-4) Face condition in rainy season: lower working efficiency for claystone in rainfall

- a-5) Mining equipment: periodical inspections and maintenance
- a-6) Mining operation: removing of the face, hard rock arrangement, troubles and damages of mining equipment, cleaning of coal floor and incleasing of belt conveyor.
- b) Basic operating hours per year
  - b-1) The number of days                      365 days ..... A
    - "                 shifts            1,095 shifts ..... B
    - "                 hours             8,760 hours ..... C
  - b-2) Weekly holidays (a day each week)
    - 365 days ÷ 7 days x 3 shifts x 8 hours
    - = 1,251 hr. .... D
  - b-3) National holidays (except Sundays)
    - 11 days x 3 shifts x 8 hours = 264 hr. ... E
  - b-4) Moslem holidays (one shift each Friday)
    - $\frac{365 - (52 + 11)}{6} \times 8 \text{ hours} = 402 \text{ hr. .... F}$
  - b-5) Schedule maintenance (1 shift each week)
    - Same as F    402 hr. .... G
  - b-6) Waiting hour due to rainfall
    - (50% operation in the 6-month rainy season from December through May)
    - Dec.-May:    182 days,
    - Weekly holidays:                                      624 hr.
    - National holidays:                                      132 hr.
    - Moslem holidays:                                      201 hr.
    - $\frac{182 \times 24 - (624 + 132 + 201 + 201)}{2} = 1,605 \text{ hr. ... H}$
  - b-7) Removing of the face and unexpected trouble
    - 20% of rest hours from D to H
    - 4,837 hr. x 0.20 = 967 hr. .... I



c) Actual operation and driving hours

Labours			Machines	
	per year		per year	
days	302	A-(D+E)	285	A-(D+E+G)
shifts	855	B-(D+E+F)	806	B-(D+E+F+G)
hours	4,275	C-(D+E+F+H+I)	3,869	C-(D+E+F+G+H+I)

iii) Total required amount of removal material per hour

a) Coal production:

$3 \times 10^6$  t/year, sp.gr. 1.28  
 $2,344 \times 10^3$  bank  $m^3$ /year  
 607 bank  $m^3$ /hour

b) Overburden removal:

Stripping ratio 2.82 bank  $m^3$ /t  
 $8,460 \times 10^3$  bank  $m^3$ /year  
 2,190 bank  $m^3$ /hour

c) Total material:

$10,800 \times 10^3$  bank  $m^3$ /year  
 2,800 bank  $m^3$ /hour

Above required amount of removal materials per hour shall be considered as required hourly minimum capacity for mining equipment design.

iv) Monthly production plan

a) Based on actual driving hours of mining equipment, monthly production plan is studied as shown on Table 7-3-1. During preparation of this table, the following operation procedures are taken into account.

a-1) Advanced overburden removal using surplus capacity of mining equipment is practised in dry season from June through November and in reverse overburden removal in the rainy season is decreased than yearly average to keep

required coal production.

a-2) Through it is one of procedures to stop coal production and to carry out overburden removal only during periodical maintenance of coal gasification plant, Table 7-3-1 is selecting continuous coal production to minimize suspension hour for removing of the face.

b) According to Table 7-3-1, required maximum actual hourly capacities for mining equipment, as a total are as follows:

b-1) Coal production : 841 bank m<sup>3</sup>/hr  
(in rainy season)

b-2) Overburden removal : 2,365 bank m<sup>3</sup>/hr  
(in dry season)

v) Major mining equipment

a) Bucket wheel excavators design

a-1) Required effective output : 2,800 bank m<sup>3</sup>/hr.

a-2) Required number of units : To mine six coal seams at maximum in five-tiered-bench faces and disperse risks of output declines resulting from such factors as unexpected mechanical troubles will require four units in operation and an additional unit reserved.  
Total; five units

a-3) Required cutting capacity per unit :  
 $2,800 \div 4 = 700$  bank m<sup>3</sup>/hr.

a-4) Nominal cutting capacity with efficiency factor taken as 0.7 ;  
 $700 \div 0.7 = 1,000$  bank m<sup>3</sup>/hr.

a-5) Required effective cutting height ... 20m  
 $100\text{m (depth)} \div 5 \text{ (benches)} = 20\text{m}$

a-6) Required ground pressure (max.) 1.0 Kg/cm<sup>2</sup>  
Based on test in Singapore involving rocks of similar properties, B.W.E. is operated at max. 1.0 Kg/cm<sup>2</sup> on the face floor in the rainy season.

a-7) Adoptable types of equipment :

Nominal cutting capacity : Soft ground

1,400 - medium hard rock 700 bank m<sup>3</sup>/hr.

Nominal cutting height; Cutting height 15m,  
digging depth 0.6m

Nominal ground pressure; 1.0 Kg/cm<sup>2</sup>

To judge cutting capacity and operating capacity of equipment when raining, the operation experiences of the Bukit Asam expansion project shall be studied in due course.

b) Belt conveyor system design

b-1) Preconditions : Pit block length

(8km ÷ 5 blocks) = 1,600m

Bench height

(100m ÷ 5 benches) = 20m

b-2) B.C. arrangement

Refer to Fig. 7-3-4 General arrangement transportation routes.

b-3) Face conveyors

Type : Shiftable belt conveyor

Length : 1,600m x 5 sets

Because a block is expected to have a service life of 6.6 years, an additional set will be purchased in the sixth year of operation to use for equipment shifting.

Capacity: 2,000 loose m<sup>3</sup>/hr. per set.

B.W.E. overburden removal capacity  
1,000 m<sup>3</sup>/hr, efficiency 70%, swell  
factor 1.4

1,000 ÷ 0.7 x 1.4 = 2,000

b-4) Cross pit conveyor

Type : Normal belt conveyor

Length : 8 sets (total length 2,800m)

Table 7-3-1 Coal Supply, Demand and Stock  
(Continuous method)

Month	Actual operating hours (h)	Coal output (1,000 t/m)	Demand (1,000 t/m)	Stock (1,000 t/m)		Required capacities		Remarks	
				Balance (1,000 t/m)	Amount (1,000 t/m)	Coal output A (m <sup>3</sup> /hr.)	Overburden removal (m <sup>3</sup> /h)	Amount of advanced overburden removal (m <sup>2</sup> /hr)	Increased amount of coal output in the rainy seasons (1,000 t/m)
1	218	232	282	-50	182	831	1,969		63
2	204	219	254	-35	147	839	1,961		60
3	218	232	282	-50	97	831	1,969		63
4	210	226	273	-47	50	841	1,959		62
5	218	232	282	-50	0	831	1,969		63
6	422	235	0	235	235	435	2,365	175	
7	437	282	235	47	282	504	2,296	106	
8	437	282	282	0	282	504	2,296	106	
9	422	273	273	0	282	505	2,295	105	
10	437	282	282	0	282	504	2,296	106	
11	422	273	273	0	282	505	2,295	105	
12	218	232	282	-50	232	831	1,969		63
Total	3,863	3,000	3,000					703	374

Bench height (m)	Length (m)	B.C. sets		
		600m BC	400 BC	200 BC
100	590	1		
80	480	1		
60	360		1	(1)*
40	250		1	(1)*
20	130			1(1)*
	Total	2	2	1(3)*

Capacity : Same as face conveyors

For equipment shifting, one 600m BC to be purchased in 6th year.

\* shows a number of BC for relays.

b-5) Trunk belt conveyors

Type : Selective charge conveyor

Length : 1,300m x 2 sets (coal x 1,  
overburden x 1)

1,000m x 1 set (overburden x 1)

Capacity: 4,000 loose m<sup>3</sup>/hr. per set

When BWEs are operated at the maximum capacity, considered are a combination of four sets for overburden removal and a set for coal mining, or that of three sets for overburden removal and 2 sets for coal mining. Assuming that each trunk B.C. has a transport capacity double that of a face B.C., two sets for overburden removal and a set for coal mining will be installed.

2,000 x 2 = 4,000 loose m<sup>3</sup>/hr.

For equipment shifting: 650m for coal, 650m for overburden removal to be purchased in the sixth year.

b-6) Overland belt conveyors

Type : Normal belt conveyor

Capacity: Same as the trunk belt conveyors

Length : For coal

1,500m x 2 sets

Additional purchases 1,600m/set x  
4 times (every six years)  
For overburden removal  
1,500m x 2 sets  
Additional purchase for shifting  
1,500m x 1 set

b-7) Spoil belt conveyor  
Type : Shiftable belt conveyor (for over-  
burden removal)  
Capacity: Same as trunk belt conveyor  
Length : 1,500m x 2 sets  
1,500m x 1 set to be additionally  
purchased

c) Overburden removal discharge

c-1) Movable spreaders  
Capacity: 4,000 loose m<sup>3</sup>/hr. x 2 sets

vi) Auxiliary mining equipment

a) Scrapers

a-1) Type : Cat 621B (tall pile 15.3, flat pile  
10.7 m<sup>3</sup>/unit)

a-2) Processing amount: (3% of a peak amount in  
June)  
2,400 m<sup>3</sup>/hr. x 3% = 72 m<sup>3</sup>/hr. in terms of  
overburden removal

a-3) Capacity

Preconditions of capacity: Transport  
distance 1,500m, loading time 2 minutes,  
discharging time a minute, speed when  
loaded 12 km/hr. = 200 m/min., speed when  
empty 30 km/hr. = 500 m/min.

Calculation of capacity: (normal rate of  
operation)

$$\frac{60 \text{ min} \times 10.7 \text{ m}^3 \times 50/60}{1500/200 + 1500/500 + 2 + 1} \times 0.6 = 24 \text{ m}^3/\text{hr}$$

a-4) Required number of units:

$$72 \div 24 = 3 \text{ units}$$

2 units for other works and pares

5 units in total

b) Shovels

b-1) Type : Cat 983 (3.9m<sup>3</sup>)  
32-ton dump loading to be possible  
at minimum

b-2) Processing amount: (7% of all materials)  
2,800 m<sup>3</sup>/hr. x 7% = 200 m<sup>3</sup>/hr.

b-3) Capacity

Preconditions of capacity:

Loading time 40 sec./cycle, empty-car  
waiting factor 0.14 (in operation 0.85),  
bucket factor 0.9

Calculation of capacity:

$$\frac{3,600 \times 3.9 \times 0.9 \times 0.85 \times 50/60 \times 0.6}{40} = 132 \text{ m}^3/\text{hr}$$

b-4) Required number of units:

$$200 \div 132 = 1.6 \text{ units} \quad 2 \text{ units}$$

c) Dump trucks

c-1) Type : Cat 769B (32-ton)

c-2) Processing amount:

Shovel processing amount (200 m<sup>3</sup>) x 0.8 =  
160m<sup>3</sup>, with others directly loaded on belt  
conveyors

c-3) Capacity

Preconditions of capacity:

Transport distance 1,500m, speed when  
loaded

15 km/hr. = 250 m/min., speed when empty

30 km/hr. = 500 m/min., loading time 5

min., dumping time a min., load 32 ÷ 1.9 =  
6.8m<sup>3</sup>/unit, loading efficiency 0.9

Calculation of capacity: (normal rate of  
operation)

$$\frac{69 \times 16.8 \times 0.9 \times 59/60}{5 + 1500/250 + 1 + 1500/500} \times 0.6 = 30 \text{ m}^3/\text{hr/unit}$$

c-4) Required number of units:

$$160 \div 30 = 5.4 \text{ units} \rightarrow 6 \text{ units}$$

d) Dozers

d-1) Type : Cat D9

d-2) Required number of units:

5 units for BWEs, 3 units for scrapers, 2  
units for shovels, 2 units for spreaders,  
2 units for others, 2 units reserved.  
16 units in total

e) Shifting tractors

e-1) Type : Cat D9 class

e-2) Required number of units:

5 units for face belt conveyors, 2 units  
for spoil belt conveyors, one unit reserved.  
8 units in total

f) Drill rigs

f-1) Type : Kawasaki KRD 11C

f-2) Processing amount: All materials  
(2,800m<sup>3</sup>/hr.) x 10%; 280m<sup>3</sup>/hr.

f-3) Capacity

Preconditions of capacity:

Drilling speed 35 m/hr., blasting pattern  
4.0 x 4.0m, bench height 20m, drilling  
length 20m

Calculation of capacity

$35 \times 50 / 60 \times 4.0 \times 4.0 \times 20 / 20 \times 0.5 = 233 \text{ m}^3/\text{hr}$

f-4) Required number of units:

$280 \div 233 = 2 \text{ sets}$

vii) Other equipment

a) Drainage systems

Main pumps : 25m<sup>3</sup>/min. x 65m, 350 kW x 32 units

Main pipes : 20" pipes with joint, 6,400m

Auxiliary drainage system, spare parts: one set

b) Water supply systems

Feed pumps : 1m<sup>3</sup>/min. x 50m, 30 kW x 2 units

Feed system : 6" pipes with joint, 5,000m

Water storage tanks: Concrete-made 500m<sup>3</sup> x 2 tanks

Spare parts: one set

c) Office and others

Office, locker room, workshop, warehouse, fuel and  
oil storage, car park, outside storage yard, spoil



dump site, sewage treatment, access road (paved with asphalt), service road (8m, 4m), diversion canal, drainage channel, settling pond.

- d) Power facility  
 Power receiving and distribution facility  
 50,000 KVA

viii) Manpower planning

a) Mine operation (actual)

BWE crew	195 person
Bench conveyor crew	150
Trunk conveyor crew	23
Coal overland BC crew	53
Spoil conveyor crew	87
Scraper crew	27
S & T crew	48
Drill crew	21
Blasting	5
General	30
Administration	13
Sub total	652

b) Maintenance

652 x 0.7 456 person

c) Sub total for workers 1,108 person

d) Administration

1,108 x 0.2 55 person

e) Total for actual 1,163 person

f) Absentee

Worker  $1,108 \times \left(\frac{1}{0.9} - 1\right) = 123$  person

Administration  $55 \times \left(\frac{1}{0.95} - 1\right) = 3$

g) Total 1,289 person

4) Conceptual plan of noncontinuous mining method

- Mobile System -

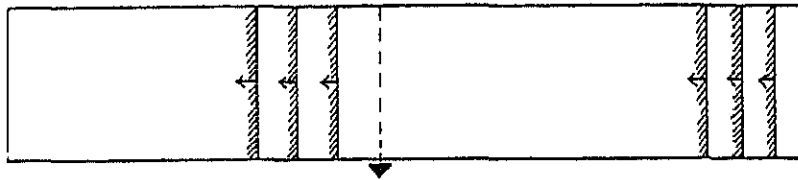
i) Mining system

a) Arrangement of faces

The strike length, an average of 7.2km, is equally divided into four blocks, each measuring 1.8km.

Mining is in practice block after block, starting from the southernmost one where coal distribution seems to be the most stable. Upon the completion of mining at a block, mining is proceeded to the adjoining one northward.

A block is divided into two parts along the direction of strike and three-tiered benches are constructed at the both sides. A block contains a total of six faces as illustrated below.



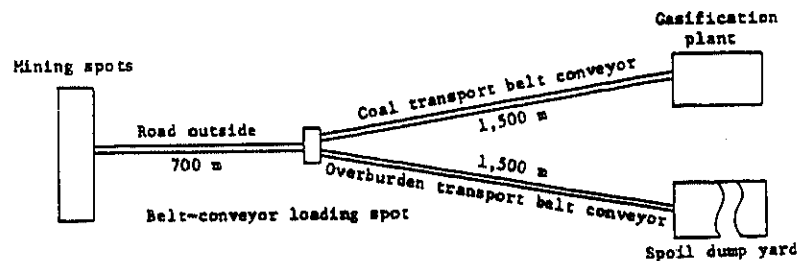
Faces are in progress toward the direction of arrows. Mining is in practice constantly at five faces, though all the six faces should be in progress on the average. As a result, transport distance within a face can be levelled (at an average of 800m). An exit of the blocks is provided at the location shown by the mark.

A bench height is designed as 10m so as to match equipment in use discussed later. Bench floor is given a rising gradient of 1/300 toward the advancing direction to provide a means of water discharge from the head of faces.

Meanwhile, the bench floor should be flat as long as the shovel dump system is employed. In addition, because of multiple-zone mining, both the coal mining and overburden removal works are conducted in the same face. These imply that changes in face conditions will be so drastic that working in a limited number of larger faces should be endangered. With these factors as well as the class of heavy machines taken into account, the design proposed here was made.

b) Transportation

A 700m-long road outside leads to a loading spot of 1,500m-long belt conveyors. The layout is illustrated below.



ii) Estimated driving hours of mining equipment  
Refer to Table 7-3-2.

3,018 hr/year.

iii) Actual amount of removal materials per hour

a) Coal production

$$3 \times 10^6 \text{ t/year} \times 1.28 \text{ bank m}^3/\text{t} = 2,344 \times 10^3 \text{ m}^3/\text{year}$$

$$2,344 \times 10^3 \text{ m}^3/\text{year} \div 3,018\text{hr}/\text{year} = 777 \text{ m}^3/\text{hr}.$$

b) Overburden removal

$$3 \times 10^6 \text{ t/year} \times 2.82 \text{ bank m}^3/\text{t} = 8,460 \times 10^3 \text{ m}^3/\text{year}$$

$$8,460 \times 10^3 \text{ m}^3/\text{year} \div 3,018\text{hr}/\text{year} = 2,803 \text{ m}^3/\text{hr}.$$

c) Total materials

$$2,344 \times 10^3 + 8,460 \times 10^3 = 10,800 \times 10^3 \text{ m}^3/\text{year}$$

$$777 + 2,803 = 3,580 \text{ m}^3/\text{hr}.$$

Above actual amount of removal materials per hour shall be considered as required minimum capacity of mining equipment design.

iv) Monthly production plan

Based on actual driving hours of mining equipment, monthly production plan is studied and shown on Table 7-3-3.

Table 7-3-2 Estimated Driving Hours of Mining Equipment

<p>1 A year in terms of hour          2 365 days x 24 hrs          3 = 8,760 hrs ... ①</p>	<p>Hours off          1,916 hrs ... ② = <math>\Sigma</math> ④ ⑥</p>	<p>④ Weekly holidays: 1,250 hrs          (365 days ÷ 7 days x 24 hrs)</p> <p>⑤ National holidays: 264 hrs          (11 days x 24 hrs)</p> <p>⑥ Moslem holidays: 402 hrs          (365 days - 52 days - 11 days)          ÷ 6 days x 8 hrs</p>	<p>⑧ Periodical repairs: 544 hrs          (17 shift x 8 hrs x 4 times)</p> <p>⑨ Waiting hours due to rainfall: 1,566 hrs ...          (182 days x 24 hrs - <math>\Sigma</math> ① ④) x <math>\frac{181}{365}</math> x 2</p>	<p>⑩ Inspection/Break: 1,184 hrs          (an hour in the rainy and 2 hours in the dry seasons per shift)</p> <p>⑪ Accidents/mechanical troubles: 531 hrs</p> <p>( ① - ② - <math>\Sigma</math> ⑧ ⑩ ) x 0.15</p>
<p>⑩ Production-suspending hours          3,829 hrs ... ⑦ = <math>\Sigma</math> ⑧ ⑩</p>	<p>Operating hours          6,844 hrs ... ③</p>	<p>⑫ Actual driving hours ... 3,018 hrs          ( ① - ② - ⑦ )</p>		

The features are as follows:

- a) Advanced overburden removal method is practised as same with continuous method
  - b) Coal production in June is suspended and all of mining equipment are applied for overburden removal, because mobile system is easy to change the face to face, resulting in no additional coal storage capacity.
  - c) By above a) and b), mining equipment capacity is retained at the average level resulting in minimized investment.
- v) Selection of major equipment (see Table 7-3-4, 7-3-5)
- a) Mining equipment  
Either mechanical or hydraulic types can be employed in carrying out mining and loading works. Assuming the use of dump trunks of the 77-ton-class, the former is selected because the hydraulic type lacks in achievement, thus failing to offer reliability.  
Calculation of the required number of units is shown in Table 7-3-4. Although, six units are possessed and each one is allocated to each face, actual operation requires five units in real terms, meaning that a unit is reserved. In addition, a bulldozer of the D-9 class and one of the D-7 class are allocated to each face. The former, designed to support mining, is installed on the upper tier of the bench, while the latter is used for bench floor arrangement.  
To prepare for bench construction, odd faces and others, a total of three hydraulic shovels, each on each level, are also allocated. Each level is also equipped with a boring machine for blasting.
  - b) Transport equipment  
Rear dumps are employed based on experience of the Bt. Asam mine using 32-ton rear dumps. In terms of class, selected are rear dumps of the 77-ton

class, which has recently been becoming popular at large mining spots. Calculation of the required number of units is shown in Table 7-3-4. To prepare for small faces, creation of faces and others, five units of the 32-ton class are also possessed. Dump transport covers up to 700m outside, then belt conveyors transport coal and overburden to a coal-consuming plant and an overburden deposit yard, respectively.

To compress overburden first piled up by spreaders, three bulldozers of the D-9 class are arranged.

c) Other equipment

Other auxiliary working machines and service equipment are shown in Table 7-3-5.

vi) Manpower planning

Administration:	192	persons
Operation:	353	"
Maintenance:	212	"
Total:	757	persons
Absentee (10%)	76	persons
Trainees (20%)	167	"
Total:	1,000	persons

(3) Preliminary Coal Mining Cost

1) Assumptions for coal mining cost

- i) Equipment and facilities cost: Based on the costs in Japan in the second quarter of 1984 and no price escalation.
- ii) Funds: 100% loan
- iii) Depreciation:
  - Period: Service life for equipment and 30 years for facilities
  - Method: Straight line method
- iv) Interest rate: 10 percent per year
- v) Labor cost: 3,300,000 Rp per year
- vi) 1 US\$ = 240 Yen = 960 Rp

2) Mining cost of continuous mining method

i) Investment

a) Initial investment	158 x 10 <sup>6</sup> US\$
b) Additional investment (up to 30th year)	128 "
Total	286 "

ii) Coal mining cost

a) Operating/labor	4.13 US\$/ton-coal
b) Depreciation	6.80 "
c) Interest	3.03 "
d) Administration	3.28 "
Total	19.70 "

3) Mining cost of non-continuous mining method

i) Investment

a) Initial investment	61 x 10 <sup>6</sup> US\$
b) Additional investment (up to 30th year)	184 "
Total	245 "

ii) Coal mining cost

a) Operation/labor	6.58 US\$/ton-coal
b) Depreciation	3.04 "
c) Interest	1.95 "
d) Administration	2.31 "
Total	13.88 "

4) Above coal mining costs are not including costs for royalty, taxes if there are, and infrastructure such as housing, hospital, school and other necessary supporting facilities as well as costs for coal transportation over 1,500m.

Also above preliminary cost estimations include some assumptions and preconditions as described in each chapter.

Therefore, needless to say, more accurate and rigid estimation is required for the feasibility.

Table 7-3-3 Coal Supply, Demand and Stock  
(Noncontinuous method)

Month	Actual operating hours (h)	Coal output (1,000 t/m)	Demand (1,000 t/m)	Stock (1,000 t/m)		Required capacities	
				Balance (1,000 t/m)	Amount (1,000 t/m)	Coal output (m <sup>3</sup> /hr.)	Overburden removal (m <sup>3</sup> /h)
1	167	282	282	0	0	1,317	2,263
2	156	254	254	0	0	1,269	2,311
3	185	282	282	0	0	1,189	2,391
4	157	273	273	0	0	1,357	2,223
5	163	282	282	0	0	1,350	2,230
6	357	0	0	0	0	0	3,580
7	326	235	235	0	0	564	3,016
8	326	282	282	0	0	674	2,906
9	357	273	273	0	0	597	2,983
10	326	282	282	0	0	675	2,905
11	314	273	273	0	0	679	2,901
12	184	282	282	0	0	1,196	2,383
Total	3,018	3,000	3,000				



Table 7-3-4 Calculation of Required Capacity and Number of Major Equipment

(1) Mining/loading equipment (Mechanical shovel)

a) Hourly capacity per unit

$$Q = \frac{3600 \times 13.0 \times 0.80 \times 1/1.30}{32} = 900 \text{ m}^3/\text{H}$$

Bucket capacity	13.0 m <sup>3</sup>
Bucket efficiency	0.8
Expansion coefficient	1/1.30
Shovel cycle time	32 sec

b) Annual capacity per unit

$$3,018 \times 900 \times 0.85 = 2,308,770 = 2,308,000 \text{ m}^3$$

Annual operating hours	3,018 H
Working efficiency	0.85

c) Required number of shovels

$$10,804,000 \div 2,308,000 \div 0.8 = 5.85 \quad 6 \text{ units}$$

Annual processing amount	Overburden	8,460,000 m <sup>3</sup>
	Coal	2,344,000 m <sup>3</sup>
	Total	10,804,000 m <sup>3</sup>

Rate of operation 0.8

(2) Transport equipment (77-ton rear dumps)

$$\text{Dump cycle time} = \frac{32}{60} \times 6 + \frac{1500}{15000/60} + \frac{1500}{20000/60}$$

$$+ 0.5 + 1 = 15.2 \text{ min.}$$

Shovel cycle time	32 sec.
Speed when loaded	15 km/H
Speed when empty	20 km/H
Waiting time	0.5 min.
Dumping time	1.0 min.

Dump actual capacity per unit =

$$\frac{60}{15.2} \times \left( \frac{18}{32} \times 77 \right) = 170.9 \text{ m}^3/\text{H}$$

Experienced actual capacity per unit

(at Bukit Asam)

$$32\text{tDT} + 18 \text{ m}^3/\text{unit}$$

b) Annual transport amount per unit

$$3,018 \times 170.9 \times 0.9 = 464,198 = 464,000 \text{ m}^3$$

Annual operating hours      3,018 H

Working efficiency            0.9

c) Required number of dump tracks

$$10,804,000 \div 464,000 \div 0.8 = 29.1 = 29 \text{ units}$$

Table 7-3-4 Equipment List  
(Noncontinuous method)

Machine	Specification	No. of req'd. Operation + Service Spare = Total ratio	Service life Hour Year	Total No. of req'd. 33	
		①	③ ④ ÷ ② x 3,018	⑤ = ① x ④	
Digging & Loading	Mechanical shovel	5 + 1 = 6	25,000	19	
	Hydraulic shovel	3	15,000	16	
	Bulldozer	9 + 2 = 11	15,000	60	
	"	5 + 1 = 6	12,000	41	
	Drill	3 + 1 = 4	16,000	18	
	Front-end loader	1	15,000	5	
	Rear dump truck	23 + 6 = 29	20,000	115	
	"	5	20,000	20	
	Belt conveyor (coal)	L-1500 m 2800 m <sup>3</sup> /h loose	1	belt 11.0	belt-3 frame-1
	" (overburden)	L-1500 m 5600 m <sup>3</sup> /h loose	1	frame 33.0	
Feeder breaker		2 + 1 = 3	4.5	13	
Spreader		1 + 1 = 2	6.0	11	
Auxiliary	Road grader	2 + 1 = 3	15,000	13	
	Road compacter	1 + 1 = 2	30,000	3	
	Fuel oil tank truck	1 + 1 = 2	"	19.9	3
	Back hoe	1 + 1 = 2	"	19.9	3
	Car for blasting	1 + 1 = 2	"	19.9	3
	Water tank truck	1 + 1 = 2	20,000	13.3	5
	Compressor	1 + 1 = 2	30,000	19.9	3
	Cleaner	1	30,000	9.9	3
	Service trailer	1 + 1 = 2	15,000	9.9	3
	Tractor	3	8	8	12
	Flat deck truck	3	5	5	20
	Patrol car	15	3	3	165
	Helicopter		1	10	3
	Bus		7	10	23
	Ambulance		1	10	3
Fire engine		2	10	6	

## 7-4 Conclusion and Recommendation

### (1) Conclusion

- 1) The No. 2 study team of JICA and BPPT at the Counterpart visited 6 relevant organizations in Jakarta, Bandung and Banko area to collect data and information as well as to discuss about the principle of the Study, that is, how to utilize effectively low grade - non transportable coal.
- 2) The applicable existing data and information on Banko area were collected in details, but those of other prospects are not yet obtained because such data are field on magnetic tape employing Shell computer system. Therefore the strategic evaluation of other prospects must be waited until the analysis of the tape.
- 3) According to the applicable exploration data, coal reserves in South Sumatra and Banko area are estimated as follows:
  - a) South Sumatra : Approximately 6 billion tons  
(estimated reserves up to 50m in depth)
  - b) Banko area : 435 million tons  
(measured resources up to 100m in depth)
- 4) According to the Shell Report and results of analysis of chip sample, quality of Banko coal is classified into brown coal with special features, such as spontaneous combustion, fragility during transportations and stock and high content of  $\text{Na}_2\text{O}$  in ash. Therefore, such unpreferable features must be adequately taken into consideration for the study of effective utilization of Banko coal.
- 5) Sampling spot and method  
It was confirmed that samples of opened outcrop are weathered and not desirable to use for coal gasification test.

For survey on coal quality in FY 1985, a small size boring machine shall be prepared to find out underground outcrops and to take up core samples from 50m in depth.

6) Preliminary mining cost estimations for two type of mining method were carried out as follows:

- a) Continuous mining method using bucket wheel excavator: Approximately 20\$/ton
- b) Non-continuous mining method using shovel and track: Approximately 14\$/ton

Above mentioned costs are not including the cost for required infrastructre, royalty and tax, etc. Therefore the actual mining cost may be increased.

7) According to the preliminary cost estimation of coal mining, non-continuous mining method seems to be superior, if methanol can be utilized as fuel instead of diesel oil.

The possibility of methanol utilization for engine shall be studied by the No. 3 study team.

(2) Recommendation

As one of the study results, the followings are recommended to carry out to make this study more effective.

1) Maps and soil data

Since there is no detailed maps for Banko area, Tanjung Enim area and Muara Enim area, preparation of maps will be required for the further study. Also some soil data should be obtained for possible plant sites to evaluate them more effectively.

2) Water recourses study

Since the discharge rate of Enim River is rather small for this project and it of Lematang River is not available, preliminary study on water resources study, such as measurement of those two rivers, availability of ground water and industrial water dam will be additionally required.

Table 4-2-7 Analytic Values of Coal Samples from Banko by Shell

Sampling Site (sampling time)	Coal Seam (m)	MOISTURE		Proximate Analytic Values (dry %)					Ash Fusion Temps (Reducing)			Ash Analysis (%)										Remarks
		(I.M.) A-D	(T.M.) A-R	ASH	V.M.	F.C.	Sulphur Kcal/kg	CV	H-G-I	I-D	H-T	P-T	SiO <sub>2</sub>	Al <sub>2</sub> O <sub>3</sub>	Fe <sub>2</sub> O <sub>3</sub>	MgO	CaO	Na <sub>2</sub> O	K <sub>2</sub> O	SO <sub>2</sub>	P <sub>2</sub> O <sub>5</sub>	
Shell Pit -BO 002 (1976)	Enim Seam Thickness* (11.4)	624	322	34	4810	4482	012	6780	612	1130	1180	1230	184	261	84	28	176	26	84	76	003	Shell Report (Nov 1976)
Shell Pit -028 (1976)	Suban-1 Seam Thickness* (5.15)	1438	317	513	4570	466	012	620	1210	1360	1390	293	304	374	438	147	024	012	493	114		
Shell Pit -BO 090 (1976)	Mangus-2 Seam Thickness* (5.0)	984	375	48	464	466	022	5587	1130	1160	1240	276	1728	183	483	167	243	040	865	092		
	Claystone Band	143	-	249	455	298	-	-	+1400	+1400	+1400	518	380	081	096	316	100	044	100	012		
N. W. Banko No.2 Pit Central Banko No.1 Pit (1977)	No.2 Suban-2 No.3 Suban-1 No.1 Mangus-2	130	311	46	404	025	5985 (gross)	59	+1400	+1400	+1400	519	356	355	033	168	038	030	626	879	Short Report on APMC Trial	
N. W. Banko (1980 ?)	NA			499	3434	3848	022	5156		1295-1315											Directorate General of Mines Analysis	

\* Thickness of the coal seam subject to sampling

Sample	Total	Ultimate Analytic Results (including water content) (Z)														Remarks
		H <sup>*</sup>	N	O <sup>*</sup>	Cl	S	Al	Ba	Ca	Fe	Mg	Na	Si	SH		
Z	140	(6843) 4481	(073) 048	(2344) 1533	002 014	(021) 014	031	001	042	022	005	005	023	10083		
-	145	(6623) 4312	(085) 062	(2375) 1588	001 013	(023) 013	063	001	042	013	011	001	053	10001		
-	140	(6569) 4083	(111) 069	(2855) 1580	001 012	(019) 012	036	001	011	011	098	005	039	10108		
Shipped to Rawang	311	627 (ad)	46 (ad)	12 (ad)										Data provided by the Indonesian Government		

## 8. Results of Survey on Banko Coal Utilization Technology

### 8-1 TECHNOLOGY FOR COAL GASIFICATION

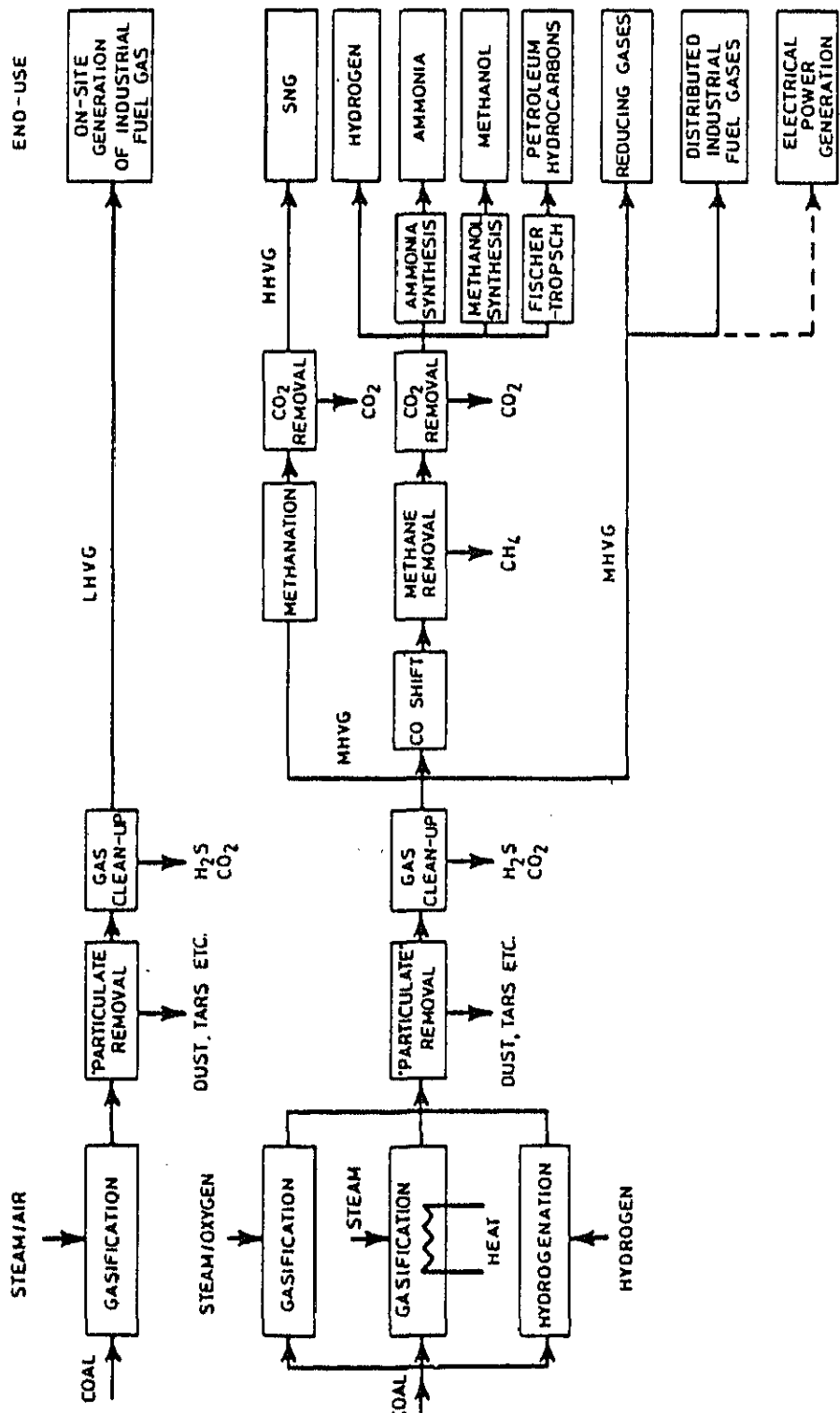
#### (1) HISTORY AND APPLICATION

Coal gasification is an old, well-proven technology that is now undergoing modernization and attracting renewed interest. The development in present day large-scale commercially proven coal gasification processes (Lurgi, Koppers-Totzek, and Winkler) took place in the 1930s and 1940s. In the late 1960s and early 1970s major installations were built in South Africa and a combination of natural gas 'shortage' and the oil 'crisis' revived serious world-wide interest in coal gasification. However, the stabilization of oil prices in 1980s, although at a price 10 times of the 1970 level, has created a change in the outlook for coal gasification. Many proposed coal gasification projects and advanced coal gasification development programs have been cancelled or postponed. However, the intensive development efforts in the 1970s have advanced some gasification processes to large-scale demonstration projects. The application of coal gasification are shown in Fig. 8-1-1. Fuel gas production is the simplest and oldest application of coal gasification. Coal-derived fuel gas is normally produced either air-blown or oxygen-blown gasifier operation. Air-blown gasification produces low heating value gas (LHVG), whereas oxygen-blown gasification produces medium heating value gas (MHVG). MHVG is rich in carbon monoxides and hydrogen (and a small portion of methane according to gasifier type), and can be used many purposes, as shown in Fig. 8-1-1, such as SNG production through methanation, methanol production, petroleum hydrocarbons production through Fischer-Tropsch process, electric power generation, etc.

#### (2) COAL GASIFICATION CHEMISTRY

In conventional combustion the carbon in the coal is essentially

Fig. 8-1-1 Application of Coal Gasification





completely converted to carbon dioxide. In contrast, gasification consists of partial combustion with a deficiency of oxygen, generally  $1/5$  to  $1/3$  of the amount theoretically required for complete combustion to carbon dioxide and water. Carbon monoxide and hydrogen, which are combustible gases, are the most predominant products of gasification. In gasification only a fraction of the carbon in the coal is oxidized completely to carbon dioxide. The heat released by partial combustion provides the bulk of the energy necessary to break chemical bonds in the coal and raise the products to reaction temperature.

As the chemistry of coal gasification is extremely complex and uncompletely defined, only a few of the important chemical reactions are discussed here.

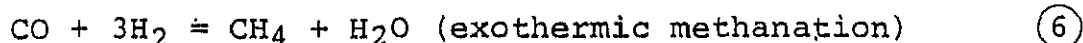
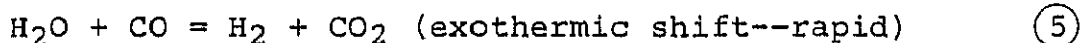
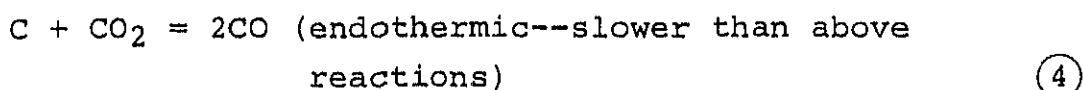
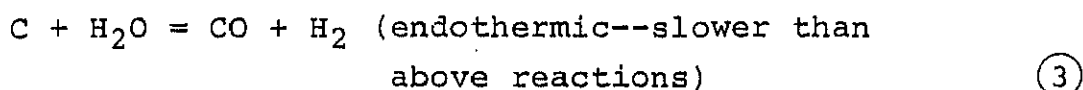
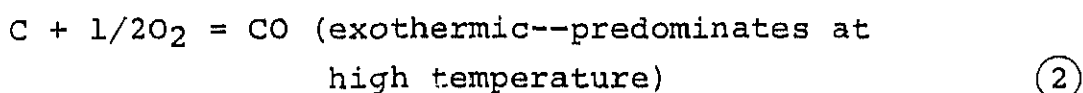
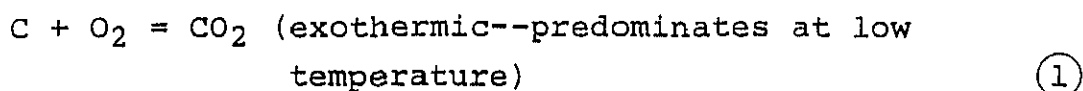
When coal is subjected to gasification conditions (i.e. substoichiometric oxygen), it is firstly dried by evaporation of the surface and inherent moisture. As the temperature rises, devolatilization occurs, and the weaker chemical bonds are broken, and tars, oils, phenols and hydrocarbon gases are formed. The rate of the devolatilization products depends strongly on the conditions under which they are released. In fixed-bed gasifiers, devolatilization products exit the gasifier with the gas because of low temperature and lack of oxygen where they are released. In fluidized-bed, entrained-flow and molten-bath gasifiers, uniform high temperatures cause cracking of the hydrocarbons to occur. Also, oxygen reacts with the devolatilization products to produce hydrogen, carbon monoxide and carbon dioxide. This happens to the greatest extent in entrained-flow and molten-bath reactors with the production almost entirely of carbon monoxide, hydrogen and carbon dioxide.

Fixed carbon (char), which remains after devolatilization, is gasified by reaction with oxygen, steam, carbon dioxide, and

hydrogen, and the gases react among themselves to produce the final gas mixture. The water gas shift reaction is favoured for control of reaction temperatures and importantly affects the H<sub>2</sub>/CO ratio but has little effect on the heating value of the product.

Fig. 8-1-2 shows important gasification reactions.

Fig. 8-1-2 PRINCIPAL COAL GASIFICATION REACTION



The first five reactions are the most prominent in the gasification systems. Reactions  $\textcircled{1}$  and  $\textcircled{2}$  are combustion of carbon which is produced by the pyrolysis of coal. Reactions  $\textcircled{3}$ ,  $\textcircled{4}$  and  $\textcircled{5}$  are called 'water-gas', 'producer-gas' and 'shift' reaction, respectively.

Water-gas reaction  $\textcircled{3}$  prefers higher temperature, and, above 1,100°C, chemical equilibrium shows carbon-monoxide and hydrogen as dominant components, while lower the reaction temperature, more carbon-dioxide. Producer gas reaction  $\textcircled{4}$  also becomes favorable at higher temperature of above 900°C.

The methanation reactions (the last two reactions) are favored by high pressures and low temperatures, but in most cases methane concentration is higher than equilibrium would predict because methane is also formed during devolatilization. Methane formation increases the thermal efficiency of gasification and the heating value of the product, while it is not preferable for production of synthesis gas.

Fig. 8-1-3 shows gas composition on equilibrium temperature at the pressure of 1 bar, 20 bar and 100 bar, respectively.

However, it is very difficult to predict conversion rate and gas composition of gasification on the basis of data of coal properties such as proximate analysis and ultimate analysis.

The rate of trace elements such as sulfur and nitrogen is also important. Sulfur in coal is converted primarily to  $H_2S$  under the reducing conditions of gasification. Approximately 5 to 15% of the sulfur is converted to  $COS$ , which adds some additional complexity to the gas cleanup step but can be handled by commercially available systems. High temperatures and low pressure favor conversion of nitrogen in coal to  $N_2$ , while the opposite conditions favor conversion of some of the nitrogen to  $NH_3$ . Small amounts of  $HCN$  are also formed. Tars, oils, and phenols, if they are not destroyed, contain some of the oxygen, nitrogen, and sulfur from the coal as more complex molecules.

### (3) CLASSIFICATION OF COAL GASIFIER

There is a large number of coal gasification processes, each having its own distinctive characteristics. However, essentially all gasifier can be classified into the four types of gasification methods: fixed-bed, fluidized-bed, entrained-flow, and molten bath gasifiers.

Fig. 8-1-4 shows the four principal types of coal gasifiers

Fig. 8-1-3 GAS COMPOSITION ON EQUILIBRIUM TEMPERATURE

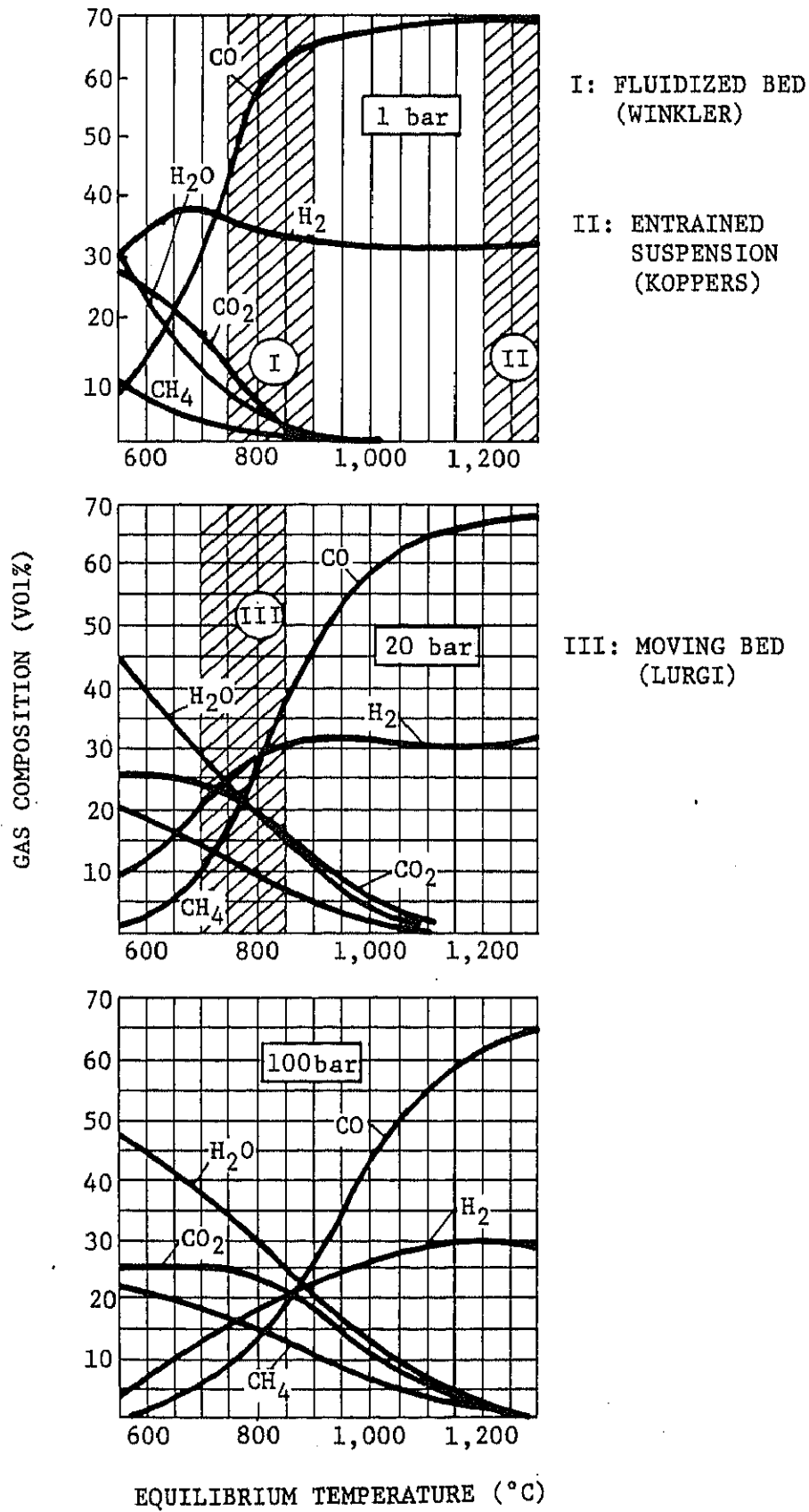
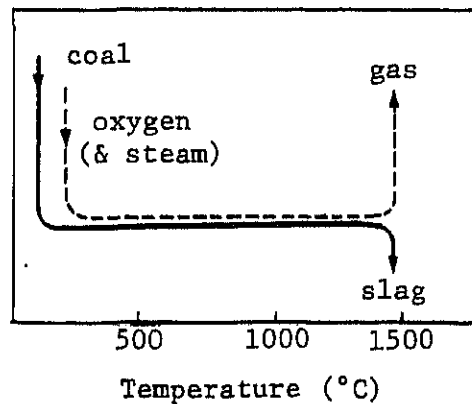
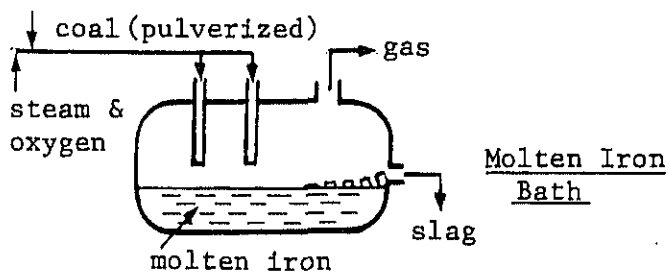
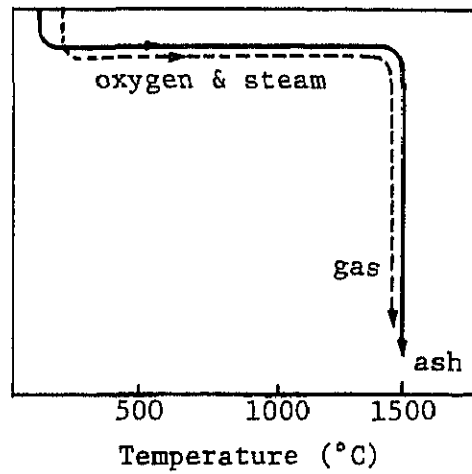
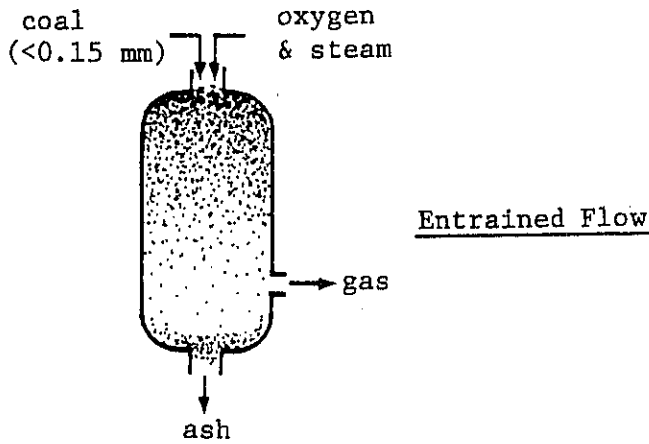
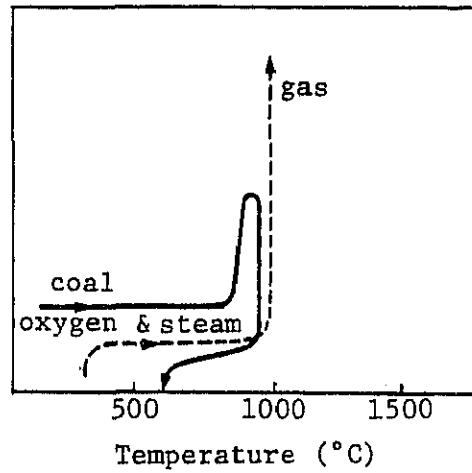
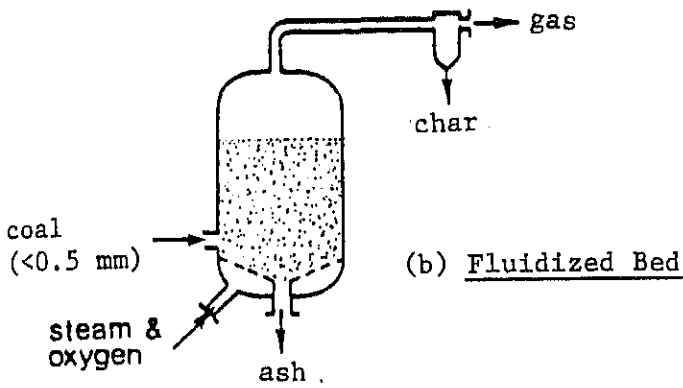
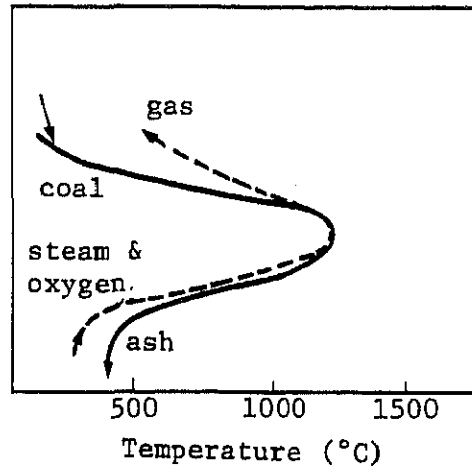
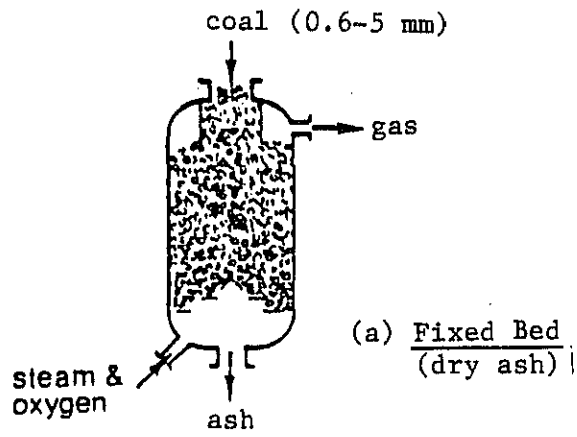


Fig. 8-1-4 Classification of Principal Coal Gasifier Types



together with temperature profiles and location of the coal, steam and oxidant (air or oxygen) inputs and the gas and ash outputs. Consideration on the types of coal gasifiers on a principle basis is useful in understanding the fundamentals of coal gasification. Table 8-1-1 summarizes the most important characteristics of each type of gasifier.

#### 1) Fixed bed gasifier

This type of reactor is also sometimes referred as a moving-bed gasifier. Lurgi gasifier is a prominent example of a fixed-bed gasifier. The process involves a series of counter-current reactions with large particles of coal moving slowly down through the bed while reacting with gases moving up through the bed. The reactions in fixed-bed gasifiers are usually described in terms of reaction zones, although there are no clear demarcations between zones. At the top, the entering coal is heated and dried in the drying zone while cooling the product gas that is about to leave the gasifier. The coal is further heated and devolatilized by hotter gas as the coal descends through the carbonization zone. Below this, the devolatilized coal is gasified by reaction with steam and carbon dioxide in the gasification zone. The highest temperatures are reached in the combustion zone near the bottom of the gasifier. There the oxygen reacts with the char which, together with ash, is all that remains of the original coal. Reaction of the char with steam under the presence of excess steam keeps the temperature of this combustion zone below the ash slagging temperature. The ash below the combustion zone is cooled by heat exchange with the entering steam and oxidant (oxygen or air).

Characteristics of fixed gasifiers include:

- a) High 'cold gas' thermal efficiency

Table 8-1-1 Important Characteristics of Principal Gasifiers

GASIFIER TYPES	FIXED BED		FLUIDIZED BED		ENTRAINED FLOW	MOLTEN IRON BATH
	Dry Ash	Slagging	Dry Ash	Agglomerating		
<u>Feed Coal Characteristics</u>						
Size	Coarse (0.6 - 5mm)	Coarse (0.6 - 5mm)	Crushed (<0.5mm)	Crushed (<0.5mm)	Pulverized (<0.15mm)	Pulverized
Acceptability of Fines	Limited	Better than dry ash	Good	Better	Unlimited	Unlimited
Acceptability of Caking Coal	Yes (with modifications)	Yes (with modification)	Possibly	Yes	Yes	Yes
Preferred Coal Rank	Low	High	Low	Any	Any	Any
<u>Operating Characteristics</u>						
Pressure (bar)	<30	24	<20	<16	40	<5 (planning)
Exit Gas Temperature (°C)	430 - 650	480	950 - 1050	950 - 1050	1250 - 1700	1400 - 1600
Oxidant Requirement	Low	Low	Moderate	Moderate	High	Moderate
Steam Requirement	High	Low	Moderate	Moderate	Low	Low or none
<u>Produced Gas Characteristics</u>						
Methane Content	High	High	Moderate	Moderate	Low	Nothing
CO <sub>2</sub> Content	High	Low	Moderate	Moderate	Moderate	Low
Tar Content	High	Moderate	Moderate	Moderate	None	None
H <sub>2</sub> S, COS	High	High	Moderate	High	High	Low
<u>Key Technical Issue</u>	Utilization of char & tar Waste water treatment		Carbon conversion Large char recycle		Raw gas cooling Char recycle	Raw gas cooling Pressurization

Source: EPRI AP-3109, with some modifications

- b) Low oxygen requirement
- c) High methane content in the produced gas
- c) Production of hydrocarbon liquids such as tars and oils
- e) Special considerations necessary to handle soft coal, caking coal and fine coal
- f) Relatively small capacity
- g) Waste water including phenol

## 2) Fluidized bed gasifier

Fluidized bed gasifier is a back-mixed reactor since feed coal particles are well mixed with coal particles already undergoing gasification. The fluidized bed is maintained at a constant temperature which is limited to being below the initial ash fusion temperature in order to avoid clinker formation and possible defluidization of the bed. Some coal particles are reduced in size during gasification and are entrained with the hot produced gas as it leaves the reactor. Ash particles are removed from below the bed and are cooled by heat exchange with the incoming steam and recycle char.

Distinguishing characteristics of fluidized bed gasifiers include:

- a) Relatively high methane content in the produced gas
- b) Large char recycle
- c) Moderate oxygen and steam requirements
- d) Special considerations necessary to handle soft coal and caking coal
- e) Production of tars and oils

This type of gasifier is preferably applied for electricity generation.

## 3) Entrained flow gasifier

Koppers-Totzek gasifier is a typical example of an entrained flow gasification reactor. The gasifier utilizes a plug



flow where the fine coal particles co-currently react with steam and oxygen. Residence time in this type of gasifier is very short. Essentially all entrained flow gasifiers use oxygen as oxidizing agent and operate at high temperature well above ash slagging conditions in order to assure reasonable carbon conversion and to provide a mechanism for ash removal.

Characteristics of entrained flow gasifiers include:

- a) Less methane content in the produced gas
- b) High temperature slagging operation
- c) Char recycle
- d) Entrainment of some molten slag in the produced gas
- e) Large amount of sensible heat in the produced gas
- f) Ability to effectively gasify a wide range of coals regardless of coal rank, coal characteristic, or the amount of coal fines.

This type of gasifier is applied for fuel and synthesis gas.

#### 4) Molten bath gasifier

Molten bath type gasifier features to hold in its bed some molten material as heat transfer medium and to feed coal and oxygen to the surface from above nozzle, or to the inside of molten material from submerged nozzle.

In the past, many types of gasifier using slag, salts and iron as molten material were studied. However, recent development is emphasized on molten slag bath type and molten iron bath type.

A molten slag bath gasifier has two reaction zones. In the lower level of gasifier, fine coal particles and oxidizing agent are blown through a non-submersion-type nozzles onto the surface of the molten slag at a high speed along a line tangent to a small circle, where devolatilization, pyrolysis and some of gasification reactions take place.

In the upper level of gasifier, char and oxidizing agent cause gasification reactions.

As obvious from above explanation, the molten slag gasifier utilizes a plug flow and the basic principle of gasifier is the same as that of an entrained blow gasifier, though molten slag acts as heat transfer medium, carbon absorber and sulfur absorber.

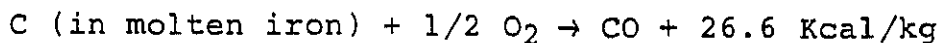
A molten iron bath gasifier utilizes molten iron which has high activity with oxygen at high temperature. Pulverized coal is blown through a non-submersion-type lance onto the surface of this molten iron at a high speed with oxidizing agent and is efficiently gasified in the molten iron bath.

The molten iron bath provides the following benefits:

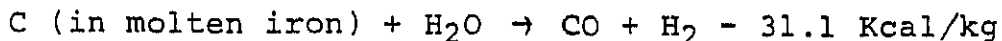
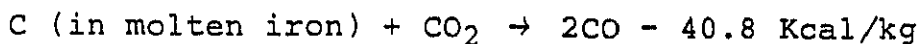
- i) The molten iron bath completely cracks the blown coal in a short period of time and not only generates hydrogen gas but dissolves and absorbs the carbon produced by cracking.
- ii) The molten iron reacts with blown oxygen and with carbon dioxide generated in the bath and becomes  $\text{FeO}$ , but this  $\text{FeO}$  is immediately reduced by carbon contained in the molten iron and becomes Fe while generating carbon monoxide gas.
- iii) Even if an excessive amount of coal is fed into the molten iron bath, the molten iron dissolves and absorbs an excessive amount of carbon preventing unoxidized carbon to escape from the gasifier.
- iv) Even if an excessive amount of oxygen is supplied, carbon contained in the molten iron bath reacts with excess oxygen preventing the generation of carbon dioxide gas.
- v) The molten iron dissolves and absorbs the sulfur contained in the coal and then transform into the molten slag.

According to above mentioned activity of the molten iron, the basic principle of coal gasification mechanism of the molten iron bath gasifier is basically different from that of other gasifiers.

It is estimated that the following reaction takes place inside of the molten bath.



In case of CO<sub>2</sub> and/or steam injection for temperature control, following reactions take place at the same time.



The basic principle of molten iron bath gasifier is already proven in steel converter of which objective is to reduce carbon content in molten raw iron.

Characteristics of molten iron bath gasifier include:

- a) No methane and tars in the produced gas
- b) Less CO<sub>2</sub> content in the produced gas
- c) Less entrainment of slag in the produced gas
- d) Large amount of sensible heat in the produced gas
- e) Ability to gasify effectively any type of coals regardless of ash properties in coal
- f) Sulfur removal by lime in gasifier

#### (4) Technology Development in Japan

After the first oil crisis in 1973, Japanese Government and many companies started the strong effort for development of coal utilization technology and at present time, 5 types of gasifier are under development.

Table 8-1-2 shows the summary of operational and performance characteristics for Japanese gasifier.

##### 1) Fluidized Bed Gasifier by CMRC

Research and development on coal gasification combined cycle power generation technology, including fluidized bed gasifier and gas clean-up system, was started in 1974 as a part of the Sunshine Project by the Ministry of International Trade and Industry (MITI) of Japan. The Coal Mining Research Counter

Table 8-1-2  
SUMMARY OF OPERATIONAL AND PERFORMANCE CHARACTERISTICS  
FOR JAPANESE GASIFIERS

Name and/or Developers	Fluidized Bed		Entrained flow		Molten Iron Bath
	CHRC EPDC NEDO	-HYBRID- EPDC NEDO	CRIEPI MHI	HITACHI	
Gasifying Medium	Air/Steam	O <sub>2</sub> /Steam	Air, O <sub>2</sub> /Steam	O <sub>2</sub> /Steam	O <sub>2</sub>
Coal Size & Feed Method					
-Size	1.5 mm	< 1 mm	200 mesh under 80-90 μ	200 mesh under 70 μ	200 mesh under 70 μ
-Feed	Lock-hopper	Resid. Oil Slurry	Pneumatic	Pneumatic	Pneumatic
Ash State	Dry	Dry	Slag	Slag	Slag
Operating Press. & Temp					
-Pressure kg/cm <sup>2</sup> G	20	30	20	9	Atm.
-Temperature °C	840-920	750-950	1000-1600	1300-1600	1400-1600
Efficiencies					
-Carbon Conversion %	94	87.7		93	> 98
-Cold Gas Efficiency %	71	71.5		70	74-80
Steam & O <sub>2</sub> Requirement					
-Steam kg/kg-coal	1.0	2.2			0.05-0.15
-O <sub>2</sub>	Air 2.1Nm <sup>3</sup>	0.5-0.7		0.8	0.5
Gas Composition Typical (vol, Dry)					
H <sub>2</sub>	14.5	31	8.4	32.3	32.7
CO	9.2	14	18.1	55.0	61.1
CH <sub>4</sub>	5.6	21	1.6	0.2	---
CO <sub>2</sub>	16.5	32	8.0	12.4	3.0
N <sub>2</sub>	53.3	--	60.7	---	---
Capacity T/D (One Unit)	40	20 (equivalent)	2	1	60
Application	Power	SNG	Power	Multi-use	Multi-use
Remarks	-Combination with Dry Desulf. System	-Coal: 4 T/D H. Oil: 8 T/D	-Two Stage entrained flow -combustor -reductor	-Two Stage Reaction	-Lime & Flux -240 T/D Pilot Plant in Sweden under construction

CHRC: Coal Mining Research Center  
EPDC: Electric Power Develop. Co.  
NEDO: New Energy Development Organization  
CRIEPI: Central Research Institute for Electric Power Industries  
MHI: Mitsubishi Heavy Industries  
CGS: Creative Gas and Steel

(CMRC) has been engaged in research and development on elemental technologies by means of pilot plants of 5 t/d and 40 t/d. According to the development plan, the development of the technology will be completed in 1987, where the New Energy Development Organization (NEDO) is carrying out the basic design of a demonstration plant, and the Electric Power Development Co., is the organization in charge of execution. The objective of the project is to develop new technology that utilizes coal in the most effective manner possible for electricity generation, while environmental protection measures also comprise extremely important objective.

Fig. 8-1-5 illustrates the conceptual structure of CMRC's fluidized bed gasifier built in Yuubari of Hokkaido, Japan, having coal processing capacity of 40 ton per day.

This gasifier features the air blown and pressurized system as well as two stage reactions including char recycle.

The air blown pressurized system possesses, aiming at power generation, such features as that it does not require a expensive air separation unit and a compressor unit of the produced gas, resulting in high thermal efficiency, as a whole.

The two stage reaction process provides better reaction conditions for each purpose. In the lower level of gasifier, char combustion takes place in oxygen rich circumstances resulting in high carbon conversion and less carbon content (0.4%) in ash, while in the upper level of gasifier devolatilization, pyrolysis and gasification in deficiency of oxygen, resulting in less tar emission and steady operation.

There is other advantage, since the gasifier operates at a temperature below the melting point of coal ash, resulting in prevention of problems caused by slagging of the ash.

Fig. 8-1-6 shows various kinds of coal ash softening temperature vs fuel ratio and illustrates that the operating

Fig. 8-1-5 Conceptual Structure of CMRC Gasifier

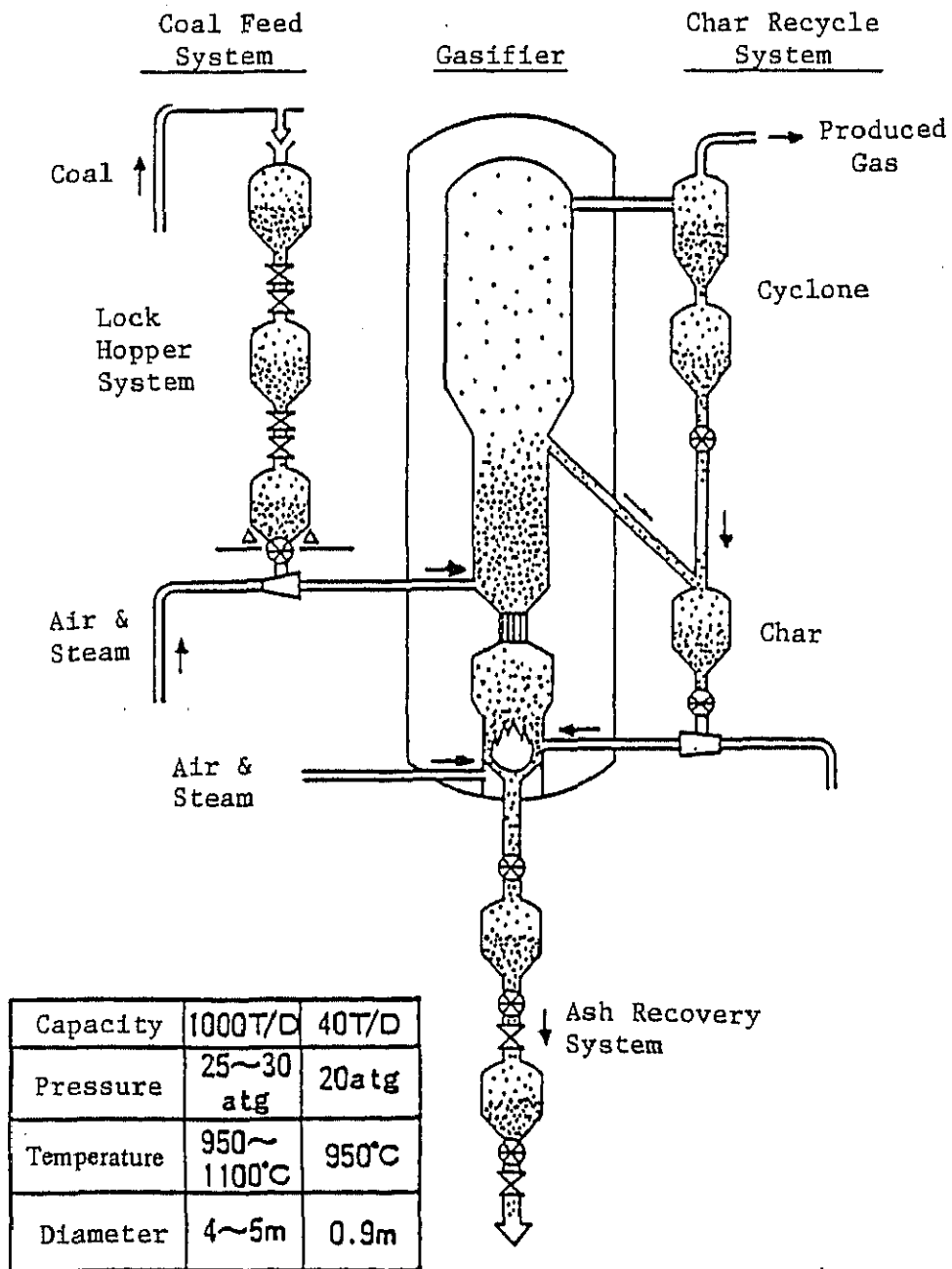
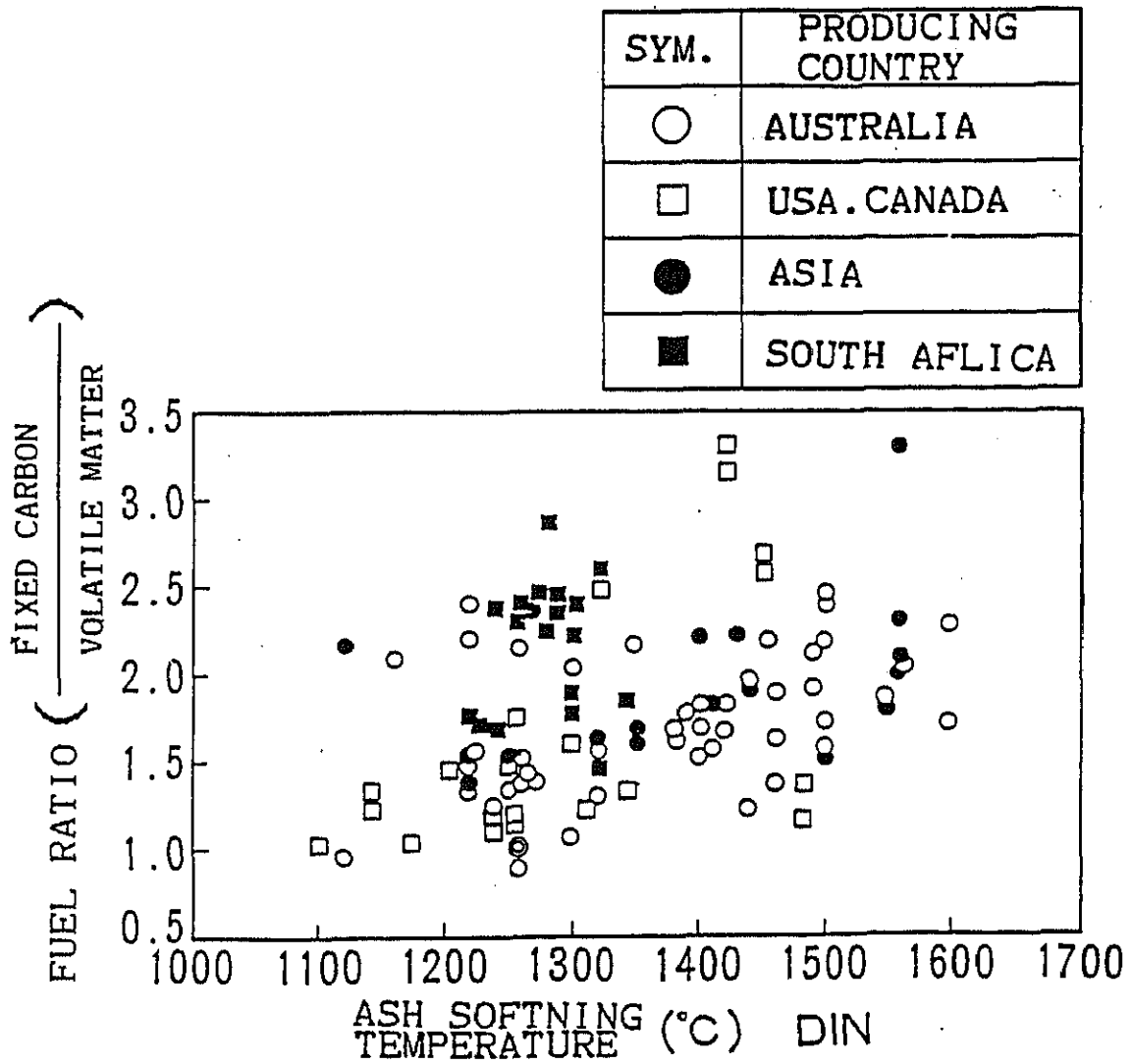


Fig. 8-1-6 VARIOUS KINDS OF COAL ASH SOFTING  
TEMPERATURE VS FUEL RATIO



temperature (1,000°C) of the lower level of gasifier is enough lower than softening temperature of coal ash.

Fig. 8-1-7 shows the results of tar emission control test, indicating that around 1,000°C of operating temperature at the upper level of gasifier can prevent tar emission.

Relation between the fluidized bed temperature and carbon conversion efficiency is shown on Fig. 8-1-8, illustrating that carbon conversion is remarkably influenced by type of coals as well as bed temperature.

More detailed data obtained by CMRC pilot plant are also shown on Table 8-1-3.

Fig. 8-1-9 shows overall flow diagram of 40 t/d coal gasification pilot plant.

The following is a brief explanation of the process.

a) Coal pretreatment

Before the raw coal is sent to the gasifier from the coal storage area, it goes through pulverizing and drying processes in order to obtain proper particle diameters and moisture content. Through these means the coal consistently conforms to a prescribed set of conditions so that gasification efficiency is also maintained at a set level, and this relates to stable operation of the power generation system as a whole.

b) Gasification

The raw coal of regulated particle size and moisture content is fed into the upper level of gasifier through the lock hopper, carbonized and gasified by the hot gas rising from the lower level of gasifier, and is converted to fuel gas with high carbon monoxide and hydrogen contents. The heat required is replenished by partial combustion of the coal. Fine



Fig. 8-1-7 RELATION BETWEEN THE FLUIDIZED  
BED TEMPERATURE AND TAR EMISSION

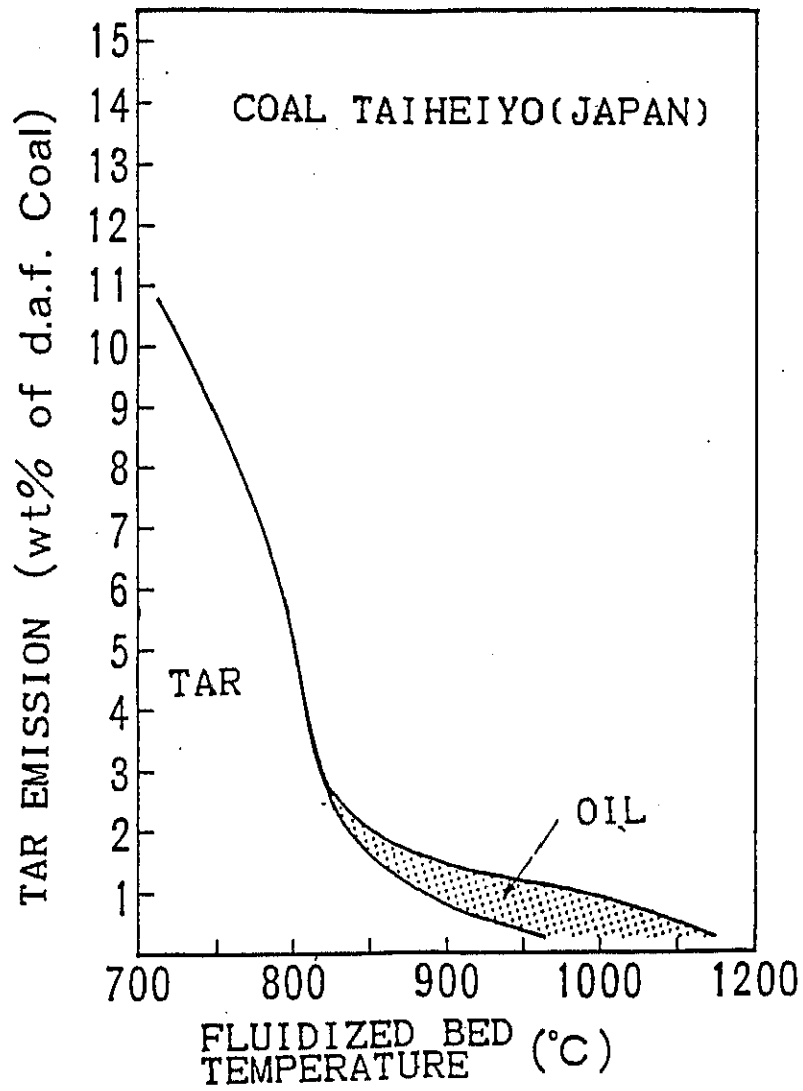


Fig. 8-1-8 RELATION BETWEEN THE FLUIDIZED BED TEMPERATURE AND CARBON CONVERSION EFFICIENCY

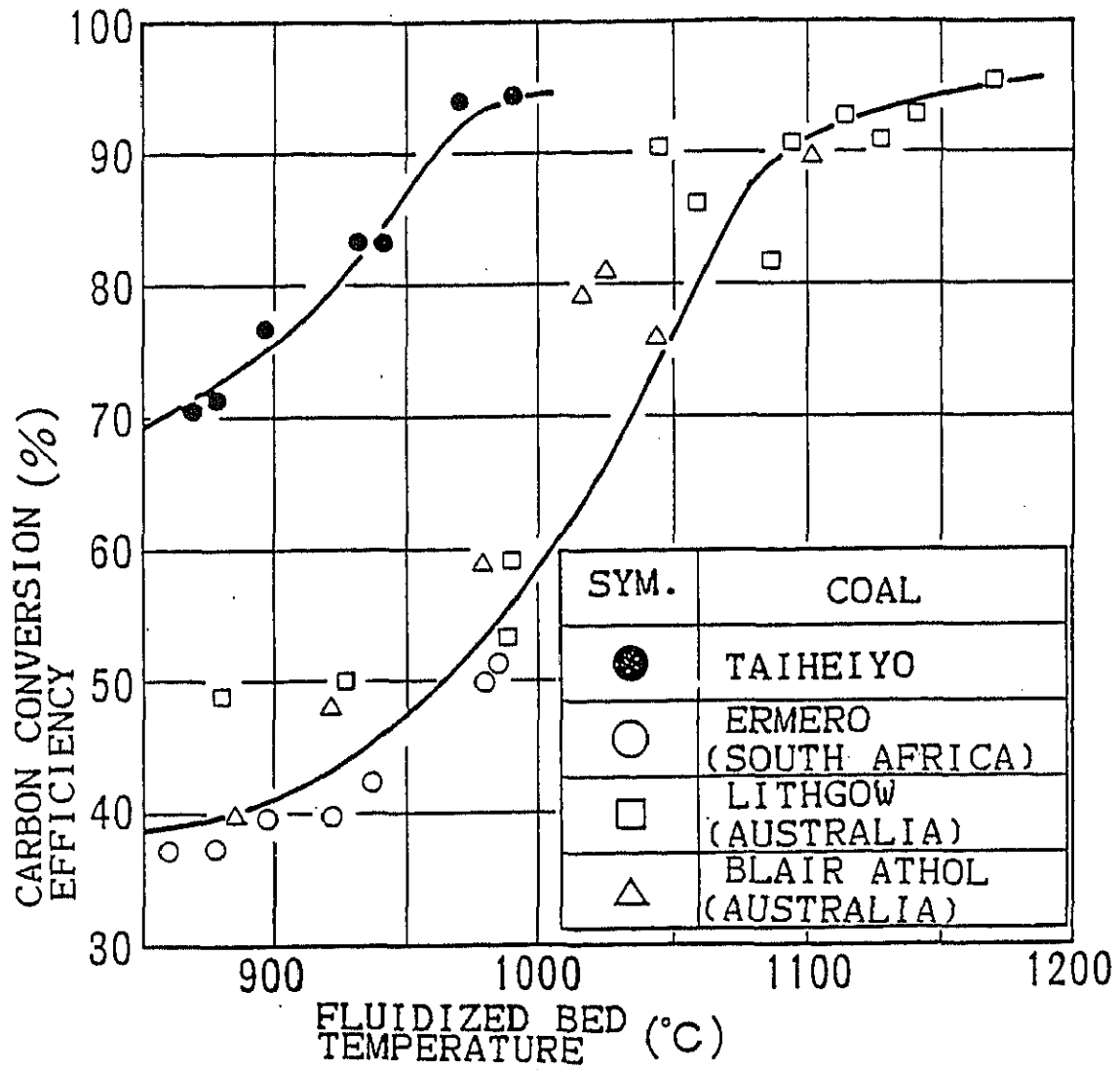


Fig. 8-1-9 OVERALL FLOW DIAGRAM FOR 40t/d FLUIDIZED  
BED COAL GASIFICATION FACILITIES

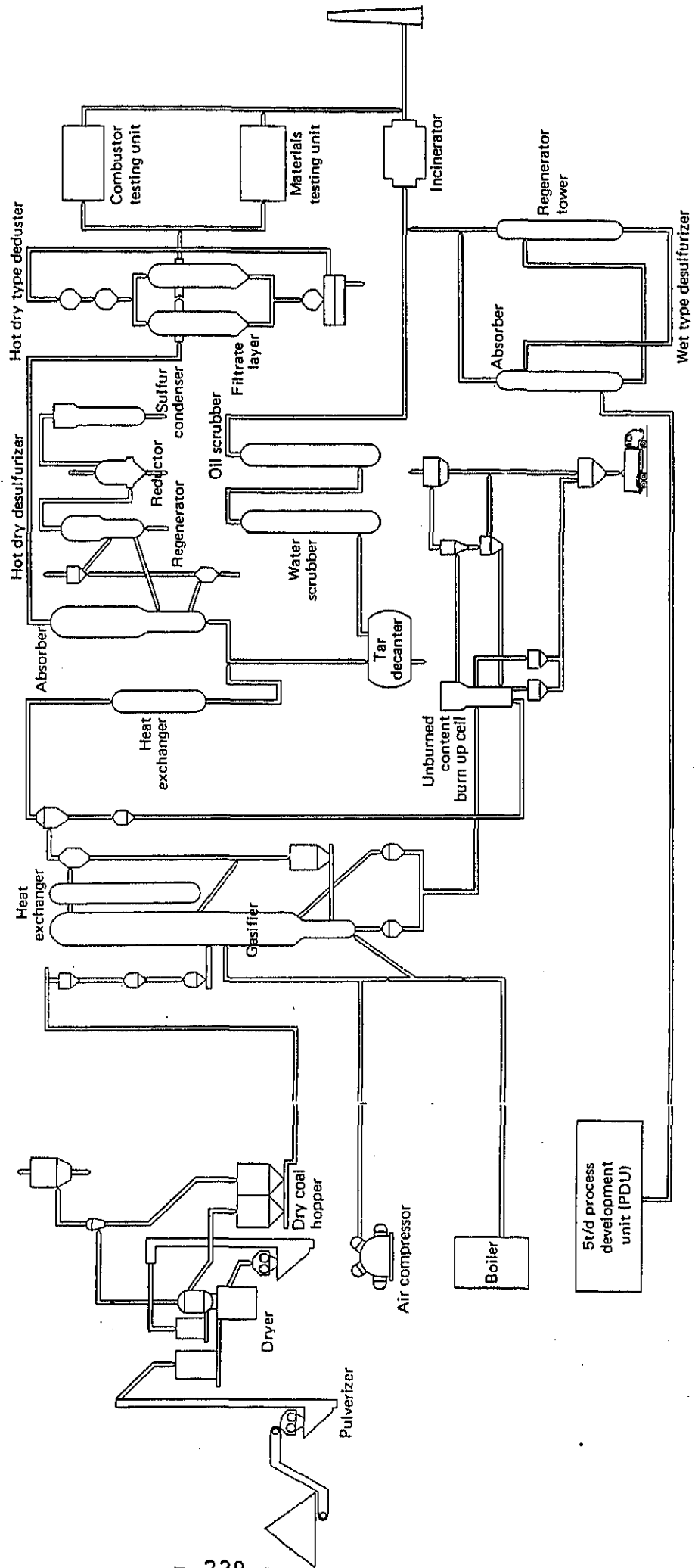


Table 8-1-3  
GASIFICATION TEST RESULTS OBTAINED  
BY CMRC PILOT PLANT

		Raw Coal	Bituminous Coal
Composition of Raw Coal	Proximate Analysis	Fixed Carbon (%)	33.7-40.0
		Ash "	14.6-17.5
		Volatile Matter "	34.2-39.0
		Inherent Moisture "	3.9-5.1
		Sulphur Content "	0.28-0.35
		Lower Heating Value (kcal/kg)	6040-6340
Operating Condition	Air/Coal Nm <sup>3</sup> /kg (%)		2.24
	Steam/Coal kg/kg "		1.03
	Pressure kg/cm <sup>2</sup> G "		20.0
	Temperature	lower level °C "	1,000
		upper level °C "	914
Gasification Test Data	Gas Composition	CO (%)	11.0
		H <sub>2</sub> "	15.8
		CO <sub>2</sub> "	15.6
		CH <sub>4</sub> "	5.4
		N <sub>2</sub> +Ar (ppm)	53.2
	Calorific Value of Product Gas (kcal/Nm <sup>3</sup> )		1325
	Desulphurization Ratio (%)		--
Gasification Efficiency "		93	
Energy Efficiency "		73	

char that is contained in the crude gas is separated in the cyclone, to return to the lower level of gasifier together with the overflow char from the upper level of gasifier, through the char hopper. Here, char combustion takes place with the air and steam, and hot gas rises to the upper level of gasifier. The ash is withdrawn by means of the lock hopper from the bottom and the overflow pipe of the lower level of gasifier..

c) Gas clean-up by hot dry type desulfurizer

The hot dry type desulfurizer under development is a fluidized bed unit that uses iron oxide as the desulfurization agent. The crude gas enters into the lowermost part of the absorber, and comes into contact with the iron oxide particles as it rises through the absorber. The sulfur compounds, which are mainly of hydrogen sulfide, are converted into iron sulfide and removed.

The desulfurized gas emerged from the top of the absorber is sent to the dust removal equipment. Meanwhile, the iron sulfide removed from the absorber is introduced to the regenerator by means of lift gas, where it is oxidized by air, converted into iron oxide and recirculated to the absorber. The sulfur oxides generated in the regenerator is taken to the reductor, reduced by coal, and recovered as element sulfur.

The absorber is operated under pressure at high temperatures ranging from 450° to 500°C, and is designed to carry out its desulfurizing function at an efficiency of more than 95%.

d) Gas clean-up by hot dry type dust removal equipment

In light of the special circumstances under which the dust is to be removed, that is under high temperature and

pressure, a granular bed type deduster is employed in this research program.

The granular bed type deduster consists of a layer of silica sand, moving at a constant speed from top to bottom, between a louver and net. The gas containing dust is filtered as it passes across the silica sand layer, and the filtrate (silica sand) is discharged from the bottom of the container by lock hopper together with the dust entrapped. After separating the trapped dust, the filtrate is recirculated and used again.

## 2) Molten Iron Bath Gasifier by SMI

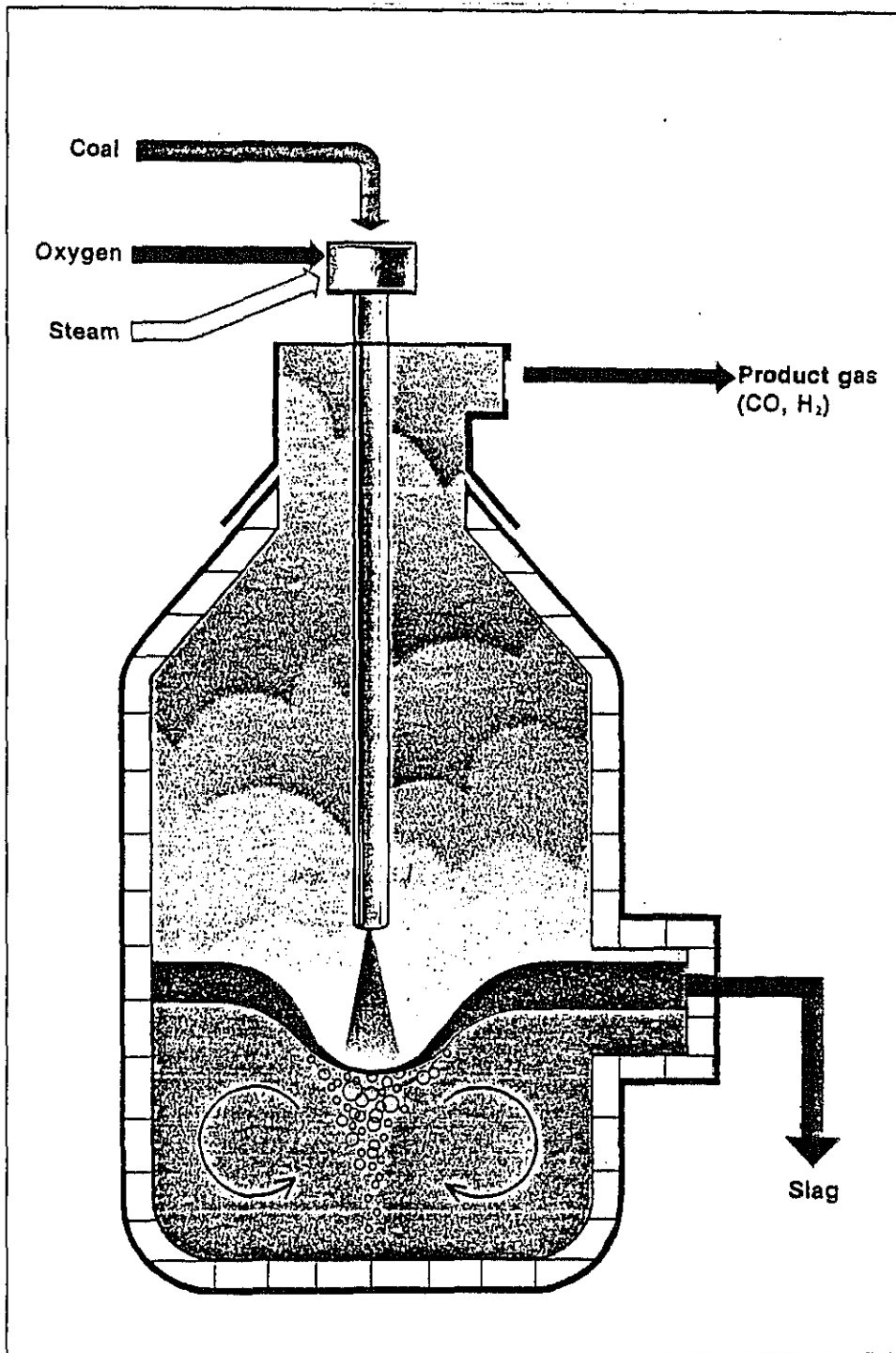
Research and development on molten iron bath gasifier was started in 1978 by Sumitomo Metal Industries Ltd., (SMI) and test operation using a large scale pilot plant, having coal processing capacity of 60 t/d, was carried out from 1980 through 1984.

The development of molten iron bath gasifier is being advanced to the demonstration operation stage. A demonstrational pilot plant of 240 t/d is under construction with joint effort of SMI and KHD (Kloekner-Humboldt - Deutz AG., West Germany) and is scheduled to start the operation in 1985.

The gasifier holds molten iron at a high temperature (1,400 - 1,600°C). When pulverized coal is blown onto this molten iron with oxygen and steam, coal is efficiently gasified in molten iron bath.

Fig. 8-1-10 illustrates principle of the molten iron bath gasifier, where molten iron acts as heat transfer medium, carbon absorber, and sulfur absorber. Some of oxygen also react with molten iron, therefore uniform reaction conditions with constant deficiency of oxygen, without any concentrated spot, can be achieved in molten iron bath. As already discussed in 8-1-(3)-4), the principle of molten iron bath gasifier created through steel production tech-

Fig. 8-1-10 Principle of the Molten Iron Bath Gasifier



nology which has long and many experiences. Therefore, major objectives of the pilot plant test were to grasp gasification characteristics of a wide variety of possible raw materials, and to obtain engineering data for commercializing it.

The pilot plant of 60 t/d was operated at temperature of 1,400 - 1,600°C and at atmospheric pressure with 10 tons of molten iron. The test results are shown on Table 8-1-4, which gives wonderful gas composition as synthesis gas, low oxygen consumption rate, high carbon conversion, high energy efficiency and high desulfurization ratio. Above table includes test data for bituminous coal, sub-bituminous coal, brown coal, petroleum pitch and residue from coal direct liquefaction.

However, the produced gas has a medium calorific value of around 2,600 kcal/Nm<sup>3</sup> and the pressure is atmospheric.

Fig. 8-1-11 shows simplified flow diagram for coal gasification plant.

The following is a brief explanation of the process.

a) Pulverizing and drying system

Coal containing moisture is fed from the coal bunker to the pulverizer through the primary crusher and dewatering drum. Hot gas is blown into the pulverizer so the coal is pulverized and dried simultaneously. The pulverized coal is then carried by the hot gas to the particle collector, and transported pneumatically from there to the coal blowing system.

b) Coal feed system

This consists of receiving tank where the pulverized coal is stored, and feed tank from which it is supplied continuously to the gasifier under pressure. The coal is transported pneumatically from the feed tank to the gasifier through the coal injection lance, using an inert gas as the carrier gas.



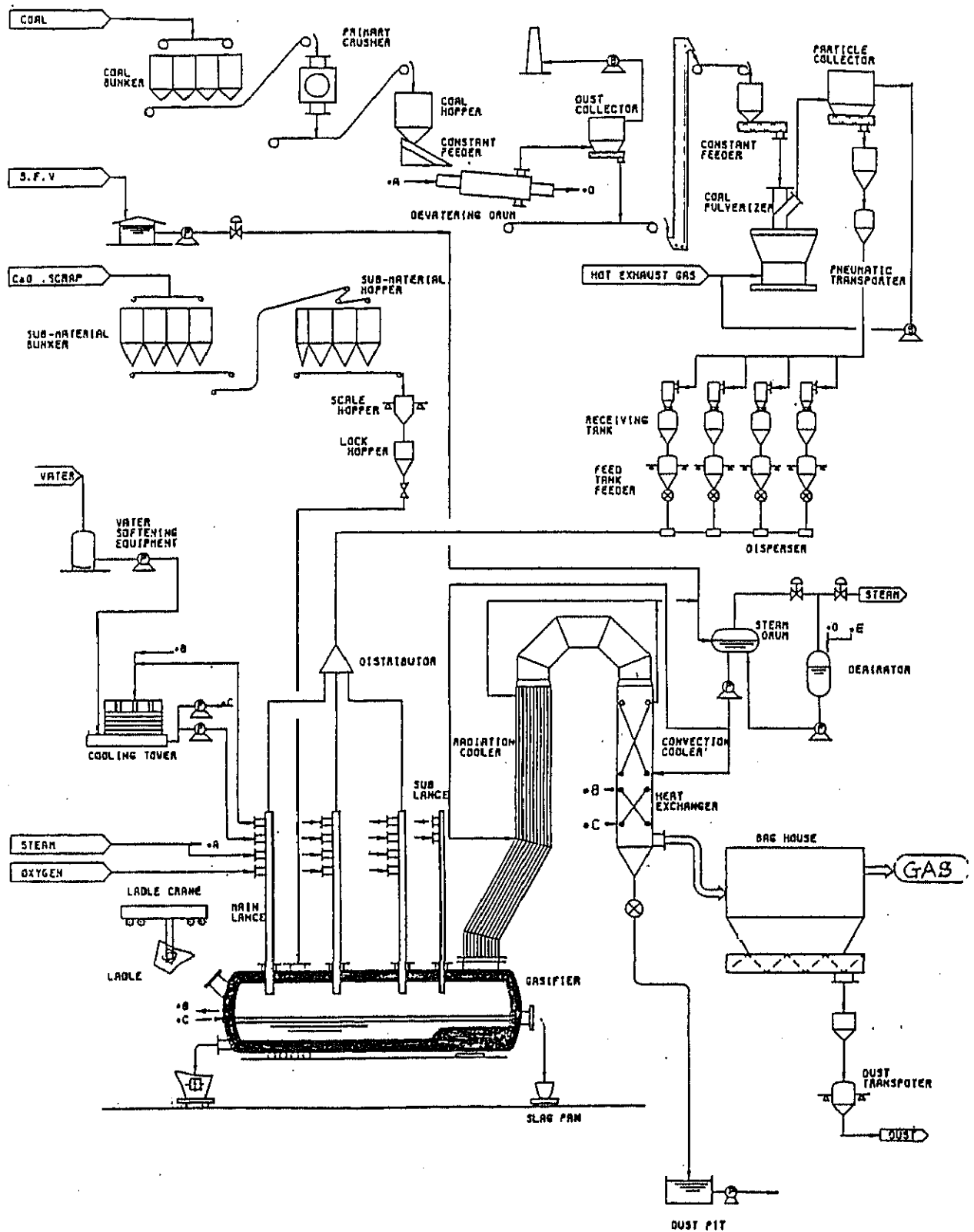
Table 8-1-4

GASIFICATION TEST RESULTS OBTAINED BY MOLTEN IRON BATH PROCESS

Composition of the Raw Material	Raw Material		Sub-bituminous coal (Australia)	Bituminous Coal (Japan)	Brown Coal (Australia)	Petroleum Pitch	Residue of Coal Liquefaction
	Analysts	Proximate					
Fixed Carbon	(%)		55.4	44.3	44.3	59.2	36.6
Ash	(%)		8.0	14.8	1.7	0.5	23.4
Volatile Matter	(%)		34.4	38.7	45.0	40.4	39.7
Inherent Moisture	(%)		2.2	2.2	9.0	0.1	0.3
Sulphur Content	(%)		0.8	0.3	0.2	5.5	3.4
Lower Heating Value	(kcal/kg)		7153	6775	5102	8438	-
Gas Composition	CO	(%)	64~66	61~64	59~63	61~65	63~66
	H <sub>2</sub>	(%)	24~26	27~29	29~30	27~32	27~30
	CO <sub>2</sub>	(%)	<3	<3	3~5	<3	1~4
	N <sub>2</sub> + Ar	(%)	<4	<4	4~5	<4	4~5
	H <sub>2</sub> S + COS	(ppm)	10~260	60~190	90~160	10~460	20~400
Calorific Value of Product Gas	(kcal/Nm <sup>3</sup> )		2640	2610	2590	2650	2680
Desulphurization Ratio	(%)		>95	>95	>95	>95	>90
Gasification Efficiency	(%)		>98	>98	>98	>98	>98
*Energy Efficiency	(%)		74.7	74.2	80.5	77.0	72.5

\* Energy efficiency: total calorific value of the product gas (kcal/kg-coal) divided by the lower heating value of the coal (kcal/kg-coal)

Fig. 8-1-11 SIMPLIFIED FLOW DIAGRAM FOR COAL GASIFICATION PLANT  
(MOLTEN IRON BATH PROCESS)



c) Gasifier

The gasifier is a simple furnace with a firebrick lining. The furnace holds several hundred tons of molten iron at temperature of 1,400 - 1,600°C. Coal, oxygen and steam are blown onto the surface of this molten iron at high speed through a specially designed non-submersion-type lances; this brings about an instantaneous gasification reaction in the iron bath.

d) Deslagging

Coal ash is transported into molten slag by the molten iron's high temperature and, having a lower specific gravity than that of the molten iron, it rises and floats on the surface of the molten iron. This molten slag is separated from the molten iron at the gasifier's deslagging zone and is then removed from the gasifier.

e) Product gas treatment system

The high temperature gas produced in the gasifier is cooled by a two-stage gas cooler where the sensible energy of gas is recovered. In addition, since the produced gas contains a small amount of dust, this dust is removed by a wet system dust washer in the second stage gas cooler and a bag house.

3) Consideration for Coal Gasification Test Plant in  
PUSPIPTEK

As discussed in 4-(3), the feasibility study on effective utilization of Banko coal includes coal gasification test to obtain basic data for the feasibility study.

The results of preliminary survey on market (6-2), and survey on Banko coal resources (6-3) play principle role of the selection of coal gasification technology to be applied to the test

plant in PUSPIPTEK. All taking into consideration, including the results of study on coal gasification technology, the candidates are evaluated as follows:

a) Field of utilization of produced gas

According to the results on preliminary survey on market, the most prospective markets of produced gas are feedstock for synthetic fuel oil, fuel methanol and urea, which are derived from synthesis gas, though some will be used as fuel for electric power generation.

As discussed in 8-1-(3) and 8-1-(4), preferable coal gasification technology for above requirement is entrained flow gasifier or molten iron bath gasifier.

Hereafter, both processes only will be discussed as candidates for coal gasification test plant.

b) Flexibility for coal quality

According to the results of survey on Banko coal quality, the coal gasification test plant must be able to gasify a wide variety of Banko coals, which feature high total moisture, low calorific value, high  $\text{Na}_2\text{O}$  content in ash and a wide range of ash fusion temperature.

Generally speaking, both of entrained flow gasifier and molten iron bath gasifier can gasify a wide range of coals, because of their high reaction temperatures and slagging ash removal systems.

At present time, no gasification process in Japan is specially designed for brown coal except one, molten iron bath gasifier, which has ever tested successfully brown coal using 60 t/d pilot plant.

Furthermore, molten iron bath gasifier has also tested successfully residue from coal direct liquefaction, which has lower calorific value and higher ash contents.

c) Operability and stability

Because the coal gasification testplant in PUSPIPTEK is small as 20 kg coal per hour, there may be some difference in test data, compared with large scale test plant.

However, the test plant is required to provide reliable and reproducible data. In this point, molten iron bath gasifier may be better because it holds molten iron which acts as big heat reservoir and reaction buffer to absorb carbon.

d) Maintenance

The test plant is required easy maintenance by using standard materials available from ordinary market.

In this point, entrained flow gasifier seems to be better.

e) Technology transfer

Process owner shall be required to dispatch operation and other engineers to Jakarta during the coal gasification testing period and also provide detailed operation manuals for counterpart.

From viewpoints of 4) and 5), it is suggested to employ one of Japan's technologies for coal gasification test plant in Indonesia.

## 8-2 Technology for Derivative Production

### (1) Technical Possibility of Synthesis Gas Utilization

Although in 1970's main raw material for processing organic chemicals was naphtha, other raw materials, such as natural gas, synthetic gas, have been highlighted due to the price escalation of naphtha and diversification of raw materials. Here synthetic gas is defined as a gas which is synthesized from coal and heavy fuel oil and contains CO and H<sub>2</sub>. Synthesis-gas-based chemistry can roughly be classified into following categories by chemical reaction.

Alcohol synthesis

Fischer-Tropsch (F/T) synthesis

Oxo-synthesis

Methanol-to-gasoline synthesis

C<sub>1</sub> chemistry

In regard to alcohol synthesis, an excellent example of synthesis-gas-based chemistry, methanol synthesis has already been industrialized, while efforts are now made for the development of fuel alcohol as well as higher alcohol to be used as chemical feedstocks.

Methanol is used in following fields;

- Substitute fuel oil (Neat or Blend, MTBE and MTG feedstock)
- Chemicals feedstock
- SCP feedstock

F/T synthesis was put into practical use for gasoline synthesis with Fe-catalysts in the 1950s, but gasoline yield is low because olefin and alcohol are simultaneously produced. To increase gasoline fraction or diesel fraction, the development of improved processes are now under way.

Oxo-synthesis is to synthesize, from synthesis gas and olefin, aldehyde of which carbon number is larger than olefin by one. This process, which can also manufacture alcohol by hydrogenation, has already been industrialized.

Employing special zeolite as a catalysis, Mobil developed a process to synthesize gasoline from methanol. The company also discovered that zeolite, used in the gasoline synthesis process, demonstrated excellent performances in aroma-synthesis and ethylene synthesis. At present, Mobil, BASF, AECI, Lurgi and Uhde are developing ethalene synthesis technologies.

Based on new catalysts technologies, such as rhodium and zeolite catalysts, involved in the aforementioned technologies, C<sub>1</sub> chemistry is now under development to produce such oxygen-containing compound as ethanol, ethylene glycol and acetic acid.

Chemicals, to be directly processed from synthesis gas, as well as other utilization of synthesis gas are illustrated in Fig. 8-2-1. Chemicals to be processed from methanol are also shown in Fig. 8-2-2. These figures are prepared in regardless of that processes have been already commercialized or under development.

Fig 8-2-1 UTILIZATION OF SYNTHETIC GAS

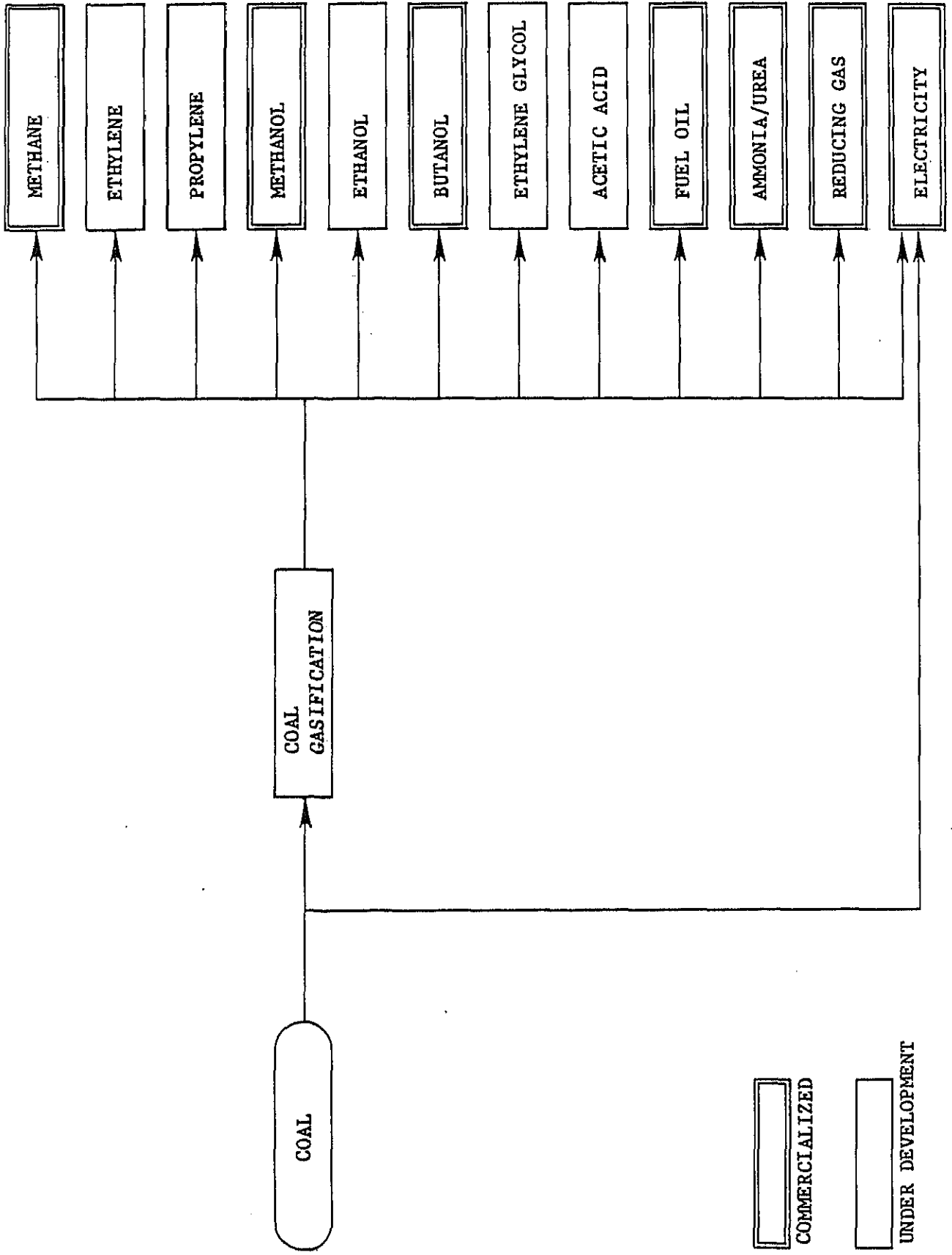
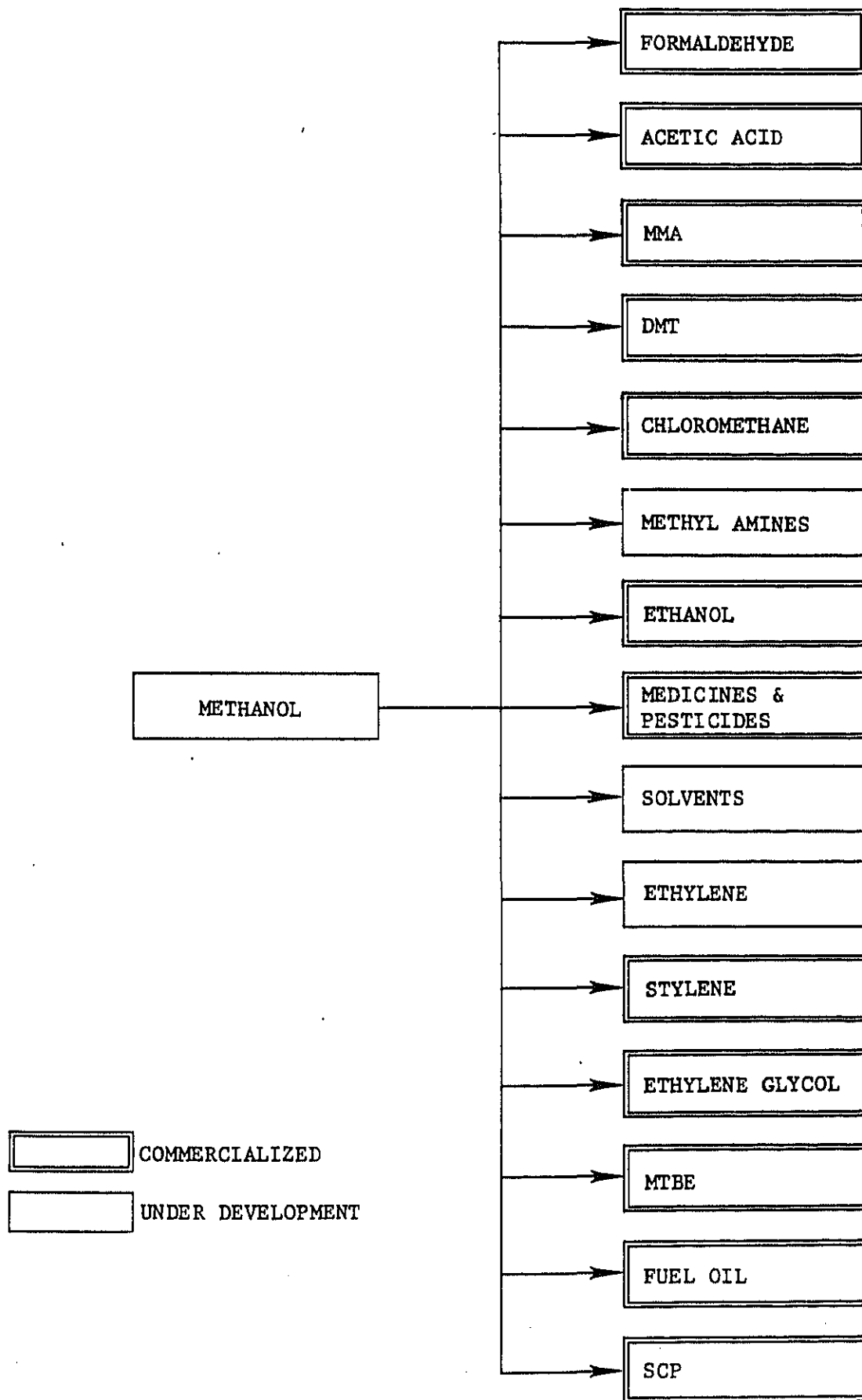




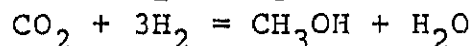
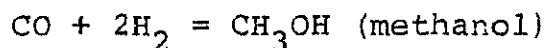
Fig 8-2-2 DERIVATIVES FROM METHANOL



## (2) Production of Methanol

### 1) Chemistry of Methanol Synthesis

Methanol can be synthesized by the catalytic reaction of synthesis gas produced by any one of commercially-available coal gasification processes which produce CO/H<sub>2</sub> mixture. Suitable gas for the catalytic production of methanol utilizing conventional zinc-chromium (high pressure) or copper-zinc-chromium (low to moderate pressures) catalysts is produced by passing the raw gas from the gasifier through a conventional water-gas shift conversion to adjust the H<sub>2</sub>:CO ratio in the gas to 2:1. After purification, this 2H<sub>2</sub>:1CO gas is sent to the catalytic reactor operated at 230 - 280°C and 50 - 150 atm, or 320 - 400°C and 200 - 400 atm (depending on the catalyst) where methanol is formed by the following reactions:



Methanol formed in the catalytic converter is condensed and unconverted gas is separated and recycled to the converter. The raw methanol is distilled for purification and higher alcohols (through pentanol) are recovered as residue together with other organic compounds.

### 2) Technology and its Development History

Methanol is currently utilized primarily as a chemical, however, interest is developing rapidly in its potential use as a fuel.

Two methanol process designs dominate the market, the Lurgi and the ICI low-pressure processes. The Lurgi process uses a tubular isothermal reactor similar to the Arge Fischer-Tropsch reactor in which tubes are packed with catalyst. As the exothermic reaction takes place in the tubes, the heat is dissipated by generating 40 atms saturated steam on the shell side of the reactor. The ICI process uses three stacked fixed beds of catalyst in which quench gas is added between beds for temperature control. Both systems are operated at pressures from 50 atms to 100 atms and use a

copper-based catalyst.

Until the mid-1960s, most methanol reactors operated at pressure of 200 atms to 340 atms and used zinc oxide catalysts promoted by chromium oxide. Improvement of sulfur clean-up processes made the use of copper catalysts and lower temperatures and pressure feasible. The new low-pressure Lurgi and ICI processes gave better economics than the high-pressure system because of low consumption of utilities, the ability to use centrifugal compressors, and higher catalyst selectivity. Essentially all of the methanol plants built since 1970 have used ICI or Lurgi low-pressure processes. Also, many of the existing high-pressure processes have been converted to these low-pressure processes.

Though both the Lurgi and ICI low-pressure methanol processes represent state-of-the-art technology, the methanol process in this report is based on the Lurgi process for consistency.

### 3) Outline of a Methanol Plant

Basis: Production of methanol

Capacity: 5,000 t/d as product methanol

No. of trains: 2

Final product: methanol 5,000 t/d

#### i) Block Flow Diagram

see Fig. 8-2-3.

#### ii) Process Description

The methanol plant covers the following process steps:

- o Dust Removal and 1st Compression
- o CO-Shift Conversion/COS Hydrolysis
- o Acid Gas Removal
- o 2nd Compression
- o Methanol Synthesis
- o Methanol Distillation

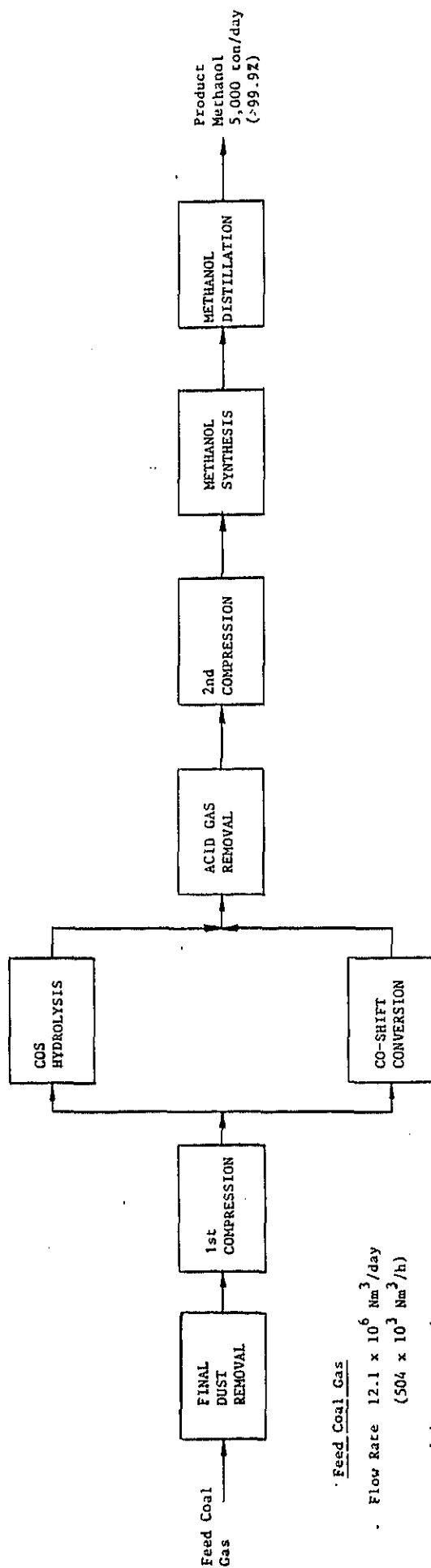
#### a) Dust Removal and 1st Compression

The raw gas leaving the gasifier at 4 kg/cm<sup>2</sup>G contains 50 mg/Nm<sup>3</sup> dust.

Fig. 8-2-3

BLOCK FLOW DIAGRAM

PRODUCTION OF METHANOL



Feed Coal Gas

Flow Rate  $12.1 \times 10^6 \text{ Nm}^3/\text{day}$   
 $(504 \times 10^3 \text{ Nm}^3/\text{h})$

Composition

H <sub>2</sub>	33.1 mol%
CO	59.6 "
CO <sub>2</sub>	3.9 "
N <sub>2</sub>	3.4 "
H <sub>2</sub> S + COS	< 200 ppm

Dry Total 100.0 mol%

Heating Value 2,818 kcal/Nm<sup>3</sup> (GHV)

Supply Condition

Pressure 4.0 kg/cm<sup>2</sup>G  
 Temperature Ambient

The dust in the raw gas is reduced to 5 mg/Nm<sup>3</sup> and this gas is compressed to 20 kg/cm<sup>2</sup>G.

b) CO-Shift Conversion/COS Hydrolysis

In order to adjust hydrogen/carbon monoxide ratio as required for the methanol production, 55% of the raw gas goes to CO-shift conversion where CO is decreased to 1% by the following reaction.



To keep the catalyst bed at proper temperature, process steam or product gas is used to dissipate the reaction heat.

Shift conversion hydrolyzes much of COS to H<sub>2</sub>S. The rest of raw gas goes to the COS hydrolysis unit where COS is also hydrolyzed into H<sub>2</sub>S.



Both gas streams are then mixed together after being cooled by generating steam, and introduced to the acid gas removal unit.

c) Acid Gas Removal

Acid gases such as H<sub>2</sub>S and CO<sub>2</sub> are to be removed from the raw gas by hot potassium carbonate (HPC) solution.



The raw gas containing about 27% of CO<sub>2</sub> and 200 ppm of H<sub>2</sub>S is fed to the absorber where CO<sub>2</sub> and H<sub>2</sub>S gases are scrubbed by HPC solution, resulting in 3.5% and a few ppm of CO<sub>2</sub> and H<sub>2</sub>S,

respectively. The treated raw gas leaving from the top of the absorber goes to the 2nd compression while the rich HPC solution from the bottom is sent to a regenerator for acid gas stripping.

d) 2nd Compression

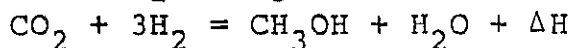
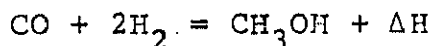
After acid gas removal, the treated raw gas is compressed to about 50 kg/cm<sup>2</sup>G.

e) Methanol Synthesis

The methanol synthesis loop comprises a recycle gas compressor, catalytic reactor, heat exchanger, condenser and separator.

The synthesis gas discharged from the compressor is fed to the catalytic reactor at about 230°C after heat exchange with the reactor effluent.

The reactor consists of multiple catalyst tubes in which the synthesis gas is converted to methanol by the following reactions:



As the catalyst cannot tolerate the adiabatic temperature rise of this exothermic reaction, the heat is dissipated by generating 40 kg/cm<sup>2</sup>G steam on the shell side of the reactor.

The reactor effluent gas at 48 kg/cm<sup>2</sup>G and 250°C is cooled by heat exchange with recycle gas and cooling medium down to ambient temperature.

The produced methanol is separated from the unconverted gas in the separator and then sent to the distillation section via a depressuring valve. The unconverted gas is recycled and mixed with the raw gas as mentioned above.

A small portion of the recycle gas is purged to limit the build-up of inerts in the loop. Purged gas is used as internal fuel for superheating of steam.

f) Methanol Distillation

The raw methanol from the synthesis unit contains a small amount of low boiling point fractions and about 5% of water. After held in the expansion vessel, the raw methanol is fed to the middle of the pretreating tower by gravity which is operated at atmospheric pressure.

In the pretreating power, dissolved gases and low boiling point fractions are stripped from the raw methanol and used as fuel. The stabilized methanol from the pretreater bottom is pumped and fed to the high pressure tower. The methanol vapor leaving the top of the tower at 120°C is cooled by heat exchange with the feed to the following atmospheric pressure tower, and then held in a reflux vessel.

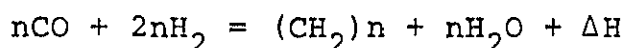
To control the liquid level in the high pressure tower, the surplus portion of the high purity methanol from the reflux vessel is sent to a storage tank through a product cooler. The methanol containing water and high boiling point components is drawn from the bottom and supplied to the atmospheric tower by gravity.

After the methanol vapor from the top of the atmospheric tower is cooled down to the ambient temperature, a half of them is sent back to the top as reflux and another half to the storage tank as products. A small portion of a mixture of high boiling point components, water, and methanol is drawn from the bottom and sent to the water treatment system in the off-site area.

### (3) Production of Synthetic Fuel Oil (F/T Synthesis)

#### 1) Chemistry of F/T Synthesis

The Production of hydrocarbons from carbon monoxide and hydrogen in the presence of Fischer-Tropsch catalyst is generally known as indirect hydrogenation.



This synthesis process was discovered in 1925 and introduced with the use of Co-catalysts into such countries as Japan and Germany in 1940s. Since then, the Lurgi-Ruhrchemie process of fixed bed type and the Kellogg process of fluidized bed type were developed, which were put to

practical use in South Africa in 1955. The only commercialized process, which can presently produce a liquid fuel from coal directly or indirectly, is the Synthol process, which is an improved version of the Kellogg process. The improved process has been commercialized in South Africa, where low-priced, and low-grade coal is abundant in order to produce gasoline fuel. Recently the Slurry F/T process, studied in the 1960s and accepting coal based synthesis gas featuring low  $H_2/CO$  ratios as feedstocks, has been gaining an increasing attention.

Table 8-2-1 shows comparisons of F/T synthesis and yields.



TABLE 8-2-1  
PRODUCT YIELDS OF F/T SYNTHESIS

Type	ARGE	Synthol
Developer	Lurgi-Ruhr Chemie	Kellogg/Sasol
Temp., °C	230	330
Pressure, atm	25	22
Conversion of (CO+H <sub>2</sub> ), %	65	85
H <sub>2</sub> : CO Ratio	1.7	2.8
Selectivity, wt%		
CH <sub>4</sub>	5.0	10.0
C <sub>2</sub> H <sub>4</sub>	0.2	4.0
C <sub>2</sub> H <sub>6</sub>	2.4	6.0
C <sub>3</sub> H <sub>6</sub>	2.0	12.0
C <sub>3</sub> H <sub>8</sub>	2.8	2.0
C <sub>4</sub> H <sub>8</sub>	3.0	8.0
C <sub>4</sub> H <sub>10</sub>	2.2	1.0
C <sub>5</sub>	3.5	8.0
C <sub>6</sub> ~ C <sub>12</sub>	19.0	31.0
C <sub>13</sub> ~ C <sub>16</sub>	15.0	5.0
C <sub>19</sub> ~ C <sub>30</sub>	23.0	4.0
C <sub>31</sub>	18.0	2.0
NAC*	3.5	6.0
Acids	0.4	1.0

(Selectivity is C-atoms converted)

\* Non acid compounds

2) Outline of F/T Synthesis Plant

Basis: Production of gasoline by Fischer Tropsch process

Capacity : Feed gas rate equivalent to  
5,000 t/d methanol production

No. of trains : 1

Final product : C<sub>5</sub>+ 6,900 BPD  
LPG 4,350 "  
Alcohol 640 "

i) Block Flow Diagram

See Fig. 8-2-4.

ii) Process Description

The F/T synthesis plant covers the following process steps:

- . Dust Removal and 1st Compression
- . CO-Shift Conversion/COS Hydrolysis
- . Acid Gas Removal
- . 2nd Compression
- . Fischer-Tropsch Synthesis
- . Product Separation
- . Hydrogen Recovery

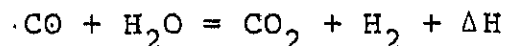
a) Dust Removal and 1st Compression

The raw gas leaving the gasifier at 4 kg/cm<sup>2</sup>G contains 50 mg/Nm<sup>3</sup> dust.

The dust in the raw gas is reduced to 5 mg/Nm<sup>3</sup> and this gas is compressed to 20 kg/cm<sup>2</sup>G.

b) CO-Shift Conversion/COS Hydrolysis

In order to adjust hydrogen/carbon monoxide ratio as required for the F/T production, 55% of the raw gas goes to CO-Shift conversion where CO is decreased to 1% by the following reaction.

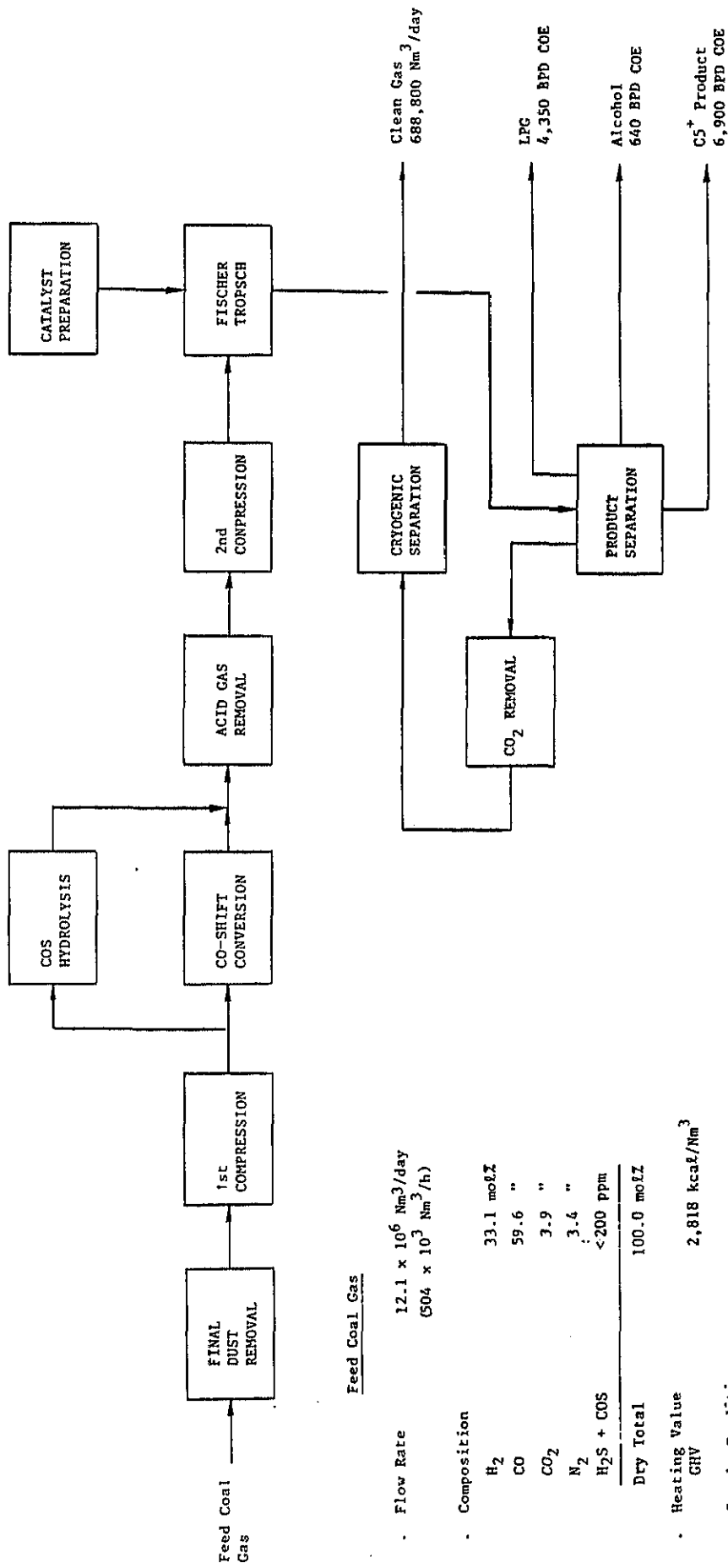


To keep the catalyst bed at a proper temperature, process steam or product gas is used to dissipate the reaction heat. Shift conversion hydrolyzes much of COS to H<sub>2</sub>S. The rest of raw gas goes to the COS hydrolysis unit where COS is also

Fig. | 8-2-4

BLOCK FLOW DIAGRAM

PRODUCTION OF SYNTHETIC FUEL OIL (F/T PROCESS)



Feed Coal Gas

Flow Rate  $12.1 \times 10^6 \text{ Nm}^3/\text{day}$   
 $(504 \times 10^3 \text{ Nm}^3/\text{h})$

Composition

H <sub>2</sub>	33.1 mol%
CO	59.6 "
CO <sub>2</sub>	3.9 "
N <sub>2</sub>	3.4 "
H <sub>2</sub> S + COS	<200 ppm
Dry Total	100.0 mol%

Heating Value  
GHV  $2,818 \text{ kcal/Nm}^3$

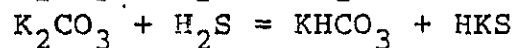
Supply Condition  
Pressure  $4.0 \text{ kg/cm}^2$   
Temperature Ambient

hydrolyzed into H<sub>2</sub>S.

Both gas streams are then mixed together after being cooled by generating steam, and introduced to the acid gas removal unit.

c) Acid Gas Removal

Acid gases such as H<sub>2</sub>S and CO<sub>2</sub> are to be removed from the raw gas by hot potassium carbonate (HPC) solution.



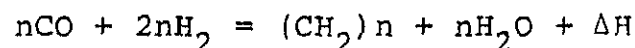
The raw gas containing about 27% of CO<sub>2</sub> and 200 ppm of H<sub>2</sub>S is fed to the absorber where CO<sub>2</sub> and H<sub>2</sub>S gases are scrubbed by HPC solution, resulting in 3.5% of CO<sub>2</sub> and a few ppm of H<sub>2</sub>S. The treated raw gas leaving the top of the absorber goes to the 2nd compression while the rich HPC solution from the bottom is sent to a regenerator for acid gas stripping.

d) 2nd Compression

After acid gas removal, the treated raw gas is compressed to about 30 kg/cm<sup>2</sup>G.

e) F/T Synthesis

After mixed with powdered catalyst, the raw gas goes upward in the F/T synthetic reactor at a temperature between 290 and 340°C where liquid products are synthesized by the general reaction:



After the reaction heat is dissipated by gas quenching and by steam generation, the reactor effluent is sent to the vertical catalyst separator and a cyclone for further separation. The converted gases from the separator go to the product separation unit, and the separated catalyst is recycled to the reactor.

f) Product Separation

The effluent gases leaving the catalyst separator is supplied to the 1st tower where heavy oil is

condensed by contacting with slurry, and then drawn off from the bottom.

After the gas from the tower top is cooled by overhead condenser, the condensates as light oil, oxygen-containing compounds and water are separated from the gas and sent to the gas-liquid separation tower. A large part of the unconverted gas from the tower top is recycled to the reaction section, and the rest is sent to the hydrogen recovery unit via acid gas treater.

In the bottom part of the gas-liquid separation tower, oil and water are separated into two phases. The oil fraction is fed to the washing tower where oxygen-containing compounds in the oil are picked up by the wash water. The water solutions from the gas-liquid separation tower and the washing tower are combined and sent to the distillation unit to recover C<sub>2</sub> - C<sub>6</sub> alcohol. The heavy oil from the 1st tower and the light oil from the washing tower are supplied to the fractionator, and LPG and various oil fractions are produced.

g) Hydrogen Recovery

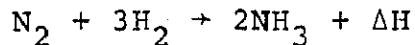
The hydrogen content of the off gas from the product separation unit is so high that hydrogen is recovered by cryogenic separation and PSA unit for reuse as a part of F/T synthesis gas. Since the off gas contains about 11% of carbon dioxide, CO<sub>2</sub> shall be reduced to several tens ppm by amine solution before being supplied to the cryogenic separation section. In the cryogenic separation section, SNG is recovered as well as hydrogen, and the rest gas is sent to fuel gas header for internal use for superheating of steam.

#### (4) Production of Ammonia and Urea

##### 1) Chemistry of Synthesis

###### i) Ammonia

All ammonia syntheses are based on the reaction:

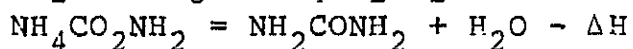
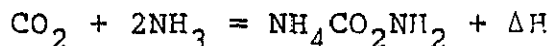


This reaction is highly exothermic, and consequently the design of the reactor must be such as to control the temperature at the point desired for the reaction deemed economical for the particular conditions chosen by the chemical engineers.

Since the volume of ammonia obtained is less than the combined volume of nitrogen and hydrogen, a pressure increase, gives a higher percentage of ammonia at equilibrium. The conversion ratio increases several folds as the pressure increases from 100 to 1,000 atm, but the ratio of ammonia in equilibrium with the reacting gases decreases continually with a temperature rise up to 1,100°C and reaches a minimum at this temperature. The commercial processes in current use are operated at about 230 kg/cm<sup>2</sup>G and 130 - 140°C at reactor inlet.

###### ii) Urea

Urea is produced from liquid ammonia and gaseous carbon dioxide at about 170 - 190°C and 134 - 145 kg/cm<sup>2</sup> abs. by the following reactions:



The first reaction is easily carried to completion, but the second usually has a conversion ratio of only 40 to 70%. Since both reactions are reversible, the equilibrium depends on the temperature, pressure, and concentration of the various components. The conversion ratio increases with rising temperature; however, urea is formed only in the liquid (solution) phase, making it necessary to maintain this phase with heat and under pressure. Because the pressure

increases rapidly with rising temperature, reaction temperatures over 210°C are rarely exceeded in commercial practice. This temperature corresponds to a conversion ratio of about 0.55. Because the combined reactions are highly exothermic, cooling is generally necessary.

2) Outline of Ammonia and Urea Plant

Basis: Production of methanol and urea

Capacity: Feed gas rate equivalent to 5,000 t/d  
methanol production

No. of trains: 2

Final product: Methanol 4,060 t/d

Urea 1,750 "

i) Block Flow Diagram

See Fig. 8-2-5.

ii) Process Description

The ammonia/urea plant consists of the following process steps:

- . Dust Removal and 1st Compression
- . CO-Shift Conversion
- . Acid Gas Removal
- . Nitrogen Washing
- . 2nd Compression
- . Ammonia Synthesis
- . NH<sub>3</sub> and CO<sub>2</sub> Compression
- . Urea Synthesis and HP Recovery
- . Low Pressure Recirculation
- . Evaporation and Prilling

a) Dust Removal and 1st Compression

The raw gas leaving the gasifier at 4 kg/cm<sup>2</sup>G contains 50 mg/Nm<sup>3</sup> dust.

The dust in the raw gas is reduced to 5 mg/Nm<sup>3</sup> and this gas is compressed to 50 kg/cm<sup>2</sup>G.

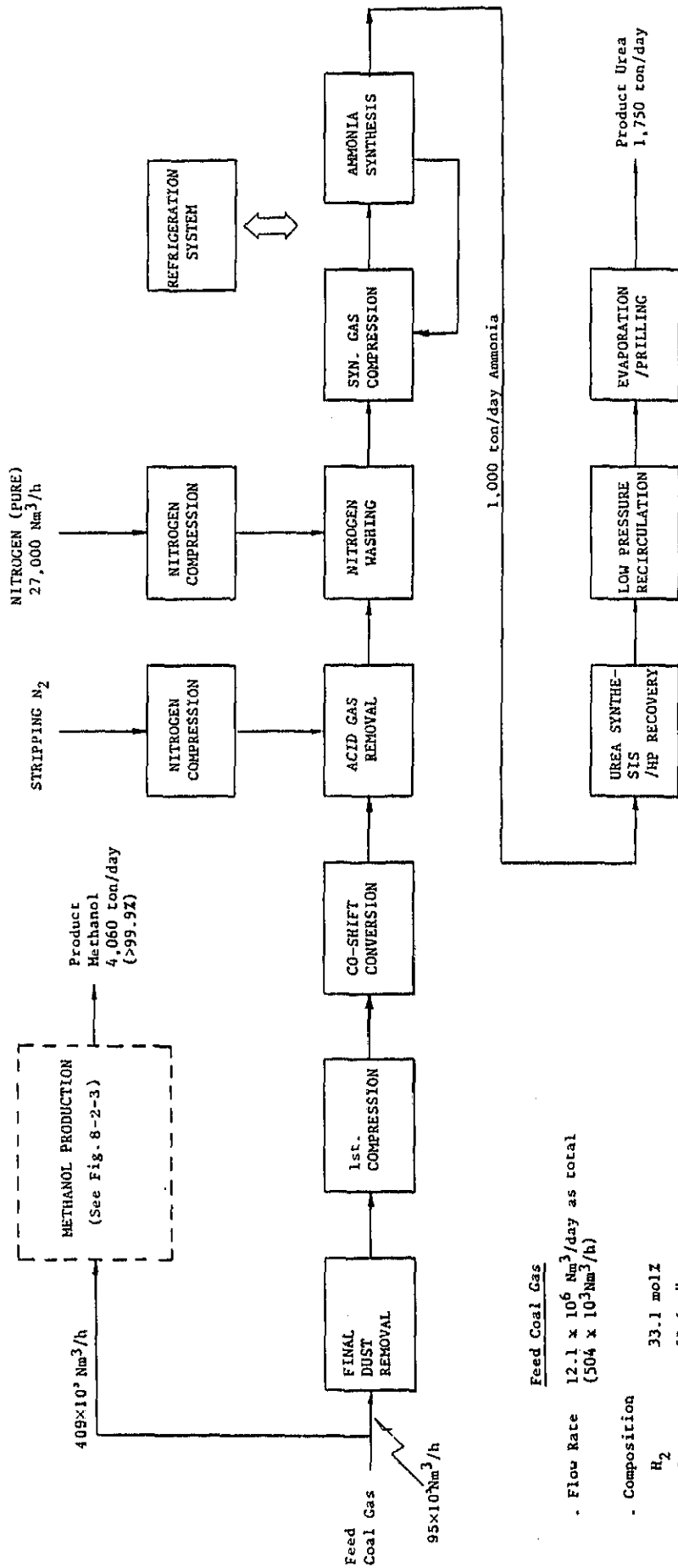
b) CO-Shift Conversion

In order to convert carbon monoxide into hydrogen by the following reaction, the raw gas goes to CO-

Fig. 8-2-5

BLOCK FLOW DIAGRAM

PRODUCTION OF AMMONIA AND UREA



Feed Coal Gas

Flow Rate 12.1 x 10<sup>6</sup> Nm<sup>3</sup>/day as total  
(50% x 10<sup>3</sup> Nm<sup>3</sup>/h)

Composition

H <sub>2</sub>	33.1 mol%
CO	59.6 "
CO <sub>2</sub>	3.9 "
N <sub>2</sub>	3.4 "
H <sub>2</sub> S+COS	<200 ppm
Dry Total	100.0 mol%

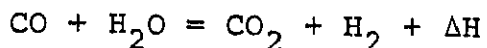
Heating Value 2,818 kcal/Nm<sup>3</sup>

Supply Condition

Pressure 4.0 kg/cm<sup>2</sup>G  
Temperature ambient



shift conversion;



To keep the catalyst bed at a proper temperature, process steam or product gas is used to dissipate the reaction heat. Shift conversion hydrolyzes also COS to H<sub>2</sub>S.

Both gas streams are then mixed together after being cooled by generating steam and introduced to the acid gas removal unit.

c) Acid Gas Removal

The Rectisol process using cold methanol as the solvent consists of an acid gas absorber, CO<sub>2</sub> stripper, H<sub>2</sub>S stripper, and distillation tower. The feed gas containing about 38% of CO<sub>2</sub> and a small amount of H<sub>2</sub>S (30 ppm) is supplied to the acid gas absorber operating at about 50 kg/cm<sup>2</sup>G and below -25°C after cooled by heat exchange with cold methanol.

In the absorber, CO<sub>2</sub> and H<sub>2</sub>S are readily scrubbed by contacting with cold methanol resulting in 10 ppm of CO<sub>2</sub> in the treated gas. The treated gas from the absorber is sent to the nitrogen washing unit at a temperature of -50°C.

The fat solvent containing CO<sub>2</sub> and a small amount of H<sub>2</sub>S is drawn from the bottom of the absorber and then sent to the CO<sub>2</sub> stripper via heat exchanger. In the CO<sub>2</sub> stripper, the solvent is depressurized to middle pressure to release the dissolved gases, and then to 1 atm to produce as pure as 99% of CO<sub>2</sub> gas which is used as the raw material for urea production.

After the remaining CO<sub>2</sub> is stripped by nitrogen, the solvent is sent to the H<sub>2</sub>S stripper where a small amount of H<sub>2</sub>S is stripped by steam. As a result, the solvent is completely regenerated and returned to the absorber after being cooled.

d) Nitrogen Washing

The gas from the acid gas removal unit still contains 10 ppm of CO<sub>2</sub> and methanol mist. Since these impurities sometimes causes plugging in the process, it is to be adsorbed by molecular sieves. The gas free from impurities is cooled and supplied to the N<sub>2</sub> washer where CO, methane, etc. are scrubbed by liquid nitrogen.

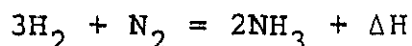
The treated gas leaving the top of the N<sub>2</sub> washer is mixed with a certain amount of nitrogen to meet the suitable H<sub>2</sub>/N<sub>2</sub> ratio for NH<sub>3</sub> synthesis. This raw gas is sent to the compression section after preheated.

e) 2nd Compression

The raw gas is compressed to about 210 kg/cm<sup>2</sup>G and introduced to the process line between heat exchangers downstream of the NH<sub>3</sub> reactor. After cooled via the heat exchangers, the raw gas is sent to the separator compressed to about 230 kg/cm<sup>2</sup>G by the gas compressor, and then sent to the NH<sub>3</sub> reactor.

f) NH<sub>3</sub> Synthesis

The feed gas is preheated to 135°C by heat exchange with reactor effluent gas and introduced to the NH<sub>3</sub> reactor. Ammonia is produced by the following reaction:



The reaction heat is dissipated by feed preheating and gas quenching. The effluent leaving the reactor at a temperature of 330°C is cooled by generating steam, feed preheating, and water cooler. The ammonia vapor is condensed in the refrigeration system, and then sent to a gas-liquid separator where product ammonia is separated from the unconverted gas. The product ammonia is pumped to storage at a temperature of -33°C via ammonia refrigeration system where

dissolved gases are released. The unconverted gas is compressed and recycled to the reactor.

g)  $\text{NH}_3$  and  $\text{CO}_2$  Compression

From the tankage the  $\text{NH}_3$  is supplied to the HP  $\text{NH}_3$  pump. The  $\text{NH}_3$  is given a pressure of 160 - 170  $\text{kg/cm}^2$  and is delivered to the HP carbamate condenser via the HP ejector.

The  $\text{CO}_2$  is supplied from acid gas removal unit to the  $\text{CO}_2$  compressor. The  $\text{CO}_2$  is given a pressure 135 - 145  $\text{kg/cm}^2$  and is delivered to the stripper. Before entering the stripper, the hydrogen removal reactor is installed. In the hydrogen removal reactor,  $\text{H}_2$  gas contained in the  $\text{CO}_2$  reacts with  $\text{O}_2$  in the air, which is introduced to the system by the process air blower, and is removed.

h) Urea synthesis and HP recovery

The synthesis stage consists of a urea reactor, a stripper for unconverted reactants, a high pressure ammonium carbamate condenser and a high pressure reactor off-gases scrubber.

Ammonium carbamate is converted in the reactor to urea and water. The heat needed for the endothermic reaction is supplied from the condensation of gaseous ammonia and carbon dioxide. In order to realize maximal urea conversion, at the stipulated optimal pressure of 140  $\text{kg/cm}^2$ , a molar proportion of ammonia to carbon dioxide in the ratio 2.85 to 1 is required in the reactor feed gas-stream.

The greater part of the unconverted carbamate is dissociated in the stripper and the ammonia and carbon dioxide are stripped off.

Low ammonia and carbon dioxide concentrations in the stripped urea solution are necessary in order to reduce the ammonia and carbon dioxide recycle, in the form of an ammonium carbamate solution, from the low pressure recirculation stage.

Condensation of ammonia and carbon dioxide gases, leaving the stripper, takes place in the high pressure carbamate condenser at synthesis pressure.

Remaining gases are condensed in the reactor and provide the heat required for dehydration of the carbamate and for heating up the reaction mixture to its equilibrium temperature.

i) Low pressure recirculation

Only one recirculation stage is required resulting from the low ammonia and carbon dioxide concentrations in the stripped urea solutions.

In this stage, remaining ammonia and carbon dioxide in the urea solution are recovered. As a result of the high efficiency of the stripper, the quantities of ammonium carbamate for recycle back to the synthesis section are also minimized, and no separate ammonia recycle is required.

j) Evaporation and prilling

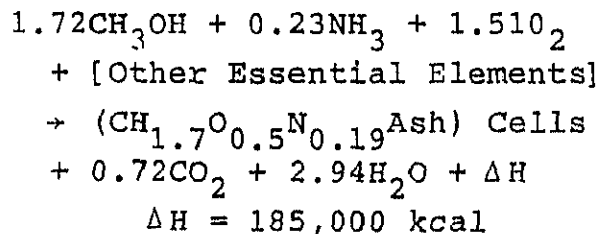
The urea solution coming from the recirculation stage contains about 72 percent by weight of urea. This solution is concentrated to 99.8 percent urea in two steps under vacuum.

The resultant molten stream is prilled with the aid of rotating prilling bucket.

(5) Production of Single Cell Protein (SCP)

1) Chemistry of SCP Synthesis

The single cell protein (SCP) is made from the bacterium *Methylophilus methylotrophus* - a gram negative aerobic rod shaped methanol obligate.



The cells are grown in a fermenter which is fed with methanol, oxygen (supplied as air), nitrogen (supplied as ammonia), and inorganic nutrients such as phosphorus, calcium, iron, magnesium and potassium.

The content of nutrients in SCP which are of key nutritional and economic importance are as follows:

Moisture	8.0 (%)
Protein	72.0
Fat	8.5
Ash	10.0
Lysine	4.1
Methionine + Cystine	1.9
Threonine	3.3
Tryptophan	1.0
Calcium	1.3
Phosphorus	2.5
Sodium	0.2

SCP contains a wide range of B vitamins and is rich in biotin, riboflavin, folic acid and pantothenic acid.

## 2) Outline of SCP Plant

Basis: A portion of product methanol is used as the feedstock for Single Cell Protein Production

Capacity: Feed gas rate equivalent to 5,000 t/d  
methanol production

No. of trains: 2

Final product: Methanol 4,500 t/d  
SCP 300 t/d

### i) Block Flow Diagram

See Fig. 8-2-6.

### ii) Process Description

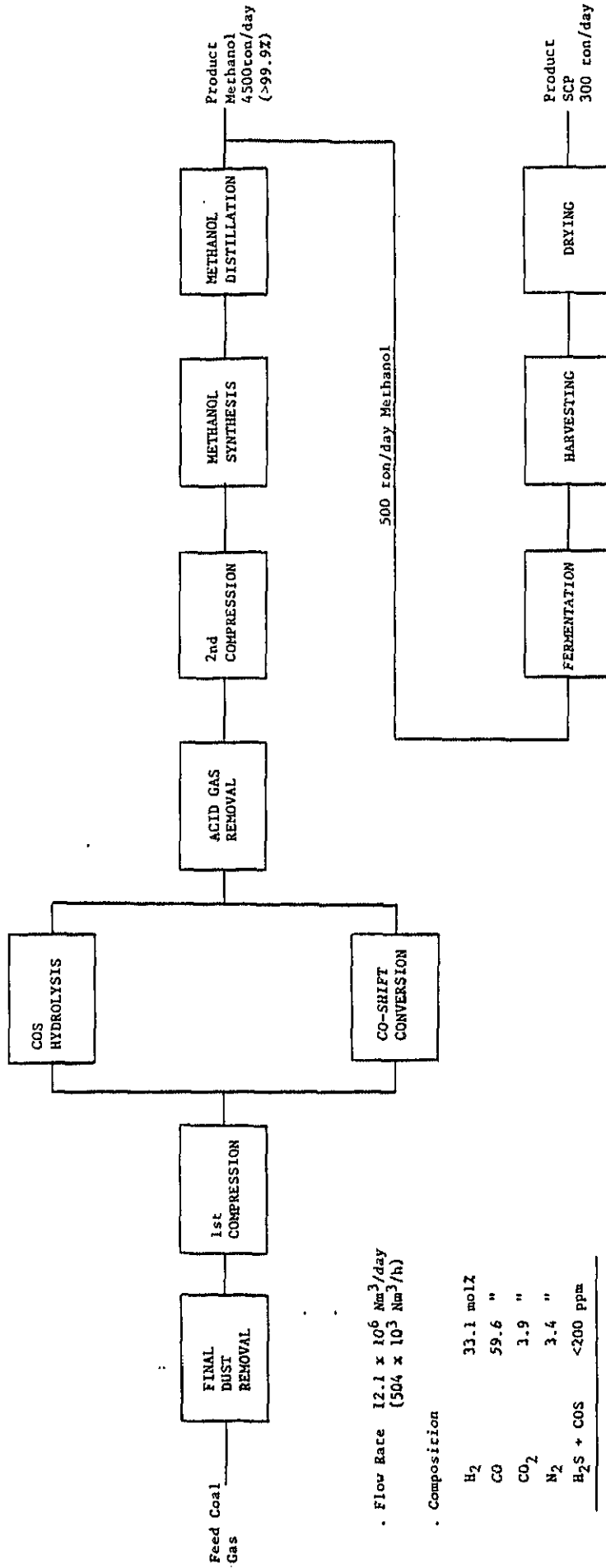
#### a) Fermentation

The heart of the plant is pressure cycle fermenter. The fermenter is a tower which has a central riser and twin downcomers. The section at the top contains the gas disengagement equipment.

Fig. 8-2-6

BLOCK FLOW DIAGRAM

PRODUCTION OF METHANOL / SINGLE CELL PROTEIN (SCP)



Flow Rate  $12.1 \times 10^6 \text{ Nm}^3/\text{day}$   
 $(504 \times 10^3 \text{ Nm}^3/\text{h})$

Composition

H <sub>2</sub>	33.1 mol%
CO	59.6 "
CO <sub>2</sub>	3.9 "
N <sub>2</sub>	3.4 "
H <sub>2</sub> S + COS	<200 ppm
Dry Total	100.0 mol%

Heating Value 2.818 kcal/Nm<sup>3</sup>  
 (GHV)

Supply Condition

Pressure 4.0 kg/cm<sup>2</sup>  
 Temperature Ambient

Compressed air, fed into the base of the riser, produces bubbles, reduces liquid density, and causes the liquid to rise. At the top, carbon dioxide is released from the solution, gases are disengaged, the liquid increases density and returns, via the downcomers, to the base of the fermenter. High circulation rates and special internal fermenter features provide good mixing, resulting in high oxygen transfer and efficient methanol utilization.

In the fermenter, the water temperature is lowered and nutrients are added to support culture growth. The fermenter is inoculated with about 20 litres of live culture.

Initial growth is rapid, the cell mass doubling every two hours. After about 30 hours from the start-up, the cell concentration reaches 3% and harvesting is commenced. Culture is withdrawn continuously from the fermenter and pumped to the harvesting section.

The liquor recycled from harvesting is balanced with added nutrients to maintain the 3% cell concentration in the fermenter.

Cooler heat exchange bundles situated near the base of the riser provide tight control of temperature.

Because the process operates with the single micro-organism, *Methylophilus methylotrophus*, absolute sterility is essential. All the nutrient feed streams are sterilised before being fed to the fermenter. Clean streams such as air, ammonia, methanol and some solutions are filtered while nutrient solutions are heat sterilized.

b) Harvesting

The culture is withdrawn continuously from the fermenter and, by addition of phosphoric acid, sulphuric acid and steam, the stream is acidified

and heated. The combined acid/heat treatment causes the cells to lyse and flocculate and produces an easily centrifuges mixture of flocs and liquor. The centrifuges produces two streams - a 20% concentrate cream which is fed to the driers and the concentrate liquor which is re-sterilized and recycled to the fermenter.

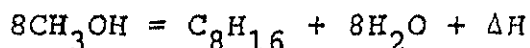
c) Drying

The concentrate cream is neutralised by the addition of sodium or calcium hydroxide. While a choice of drying systems is possible, the SCP plant uses usually a flash drier to produce a granular product. The flash drier comprises a furnace, an air lift tube, a product collection cyclone, a granulator and a gas scrubbing system. Recycled granules and fines are mixed with concentrate cream in the granulator and the resultant wet granules are fed into the base of the air lift tube. Hot furnace gases transport the granules up the tube and the moisture is rapidly flashed off. The granules are captured in the cyclone; some are returned to the granulator and the remainder pass through a fluid bed cooler to bulk storage bunkers.

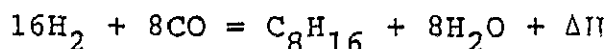
(6) Production of Gasoline from Methanol

1) Chemistry of Reaction

The MTG process which produces gasoline from methanol can be described by the following generalized reaction:



In terms of original synthesis gas the generalized reaction is:



Mobil developed the MTG process in the mid-1970s, and the development appears to be related to catalyst work they were conducting to improve Fischer-Tropsch selectivity.



This development is based on a zeolite catalyst that results in high selectivity of molecular weight of products by closely controlling the catalyst pore diameter. The advantage of this reaction over Fischer-Tropsch is a higher selectivity to desired products. Furthermore, the octane number of the raw MTG gasoline is higher than that of the gasoline range liquids from the Fischer-Tropsch reaction.

2) Outline of Methanol-To-Gasoline (MTG) Plant

Basis: Production of gasoline via methanol

Capacity: Feed gas rate equivalent to 5,000 t/d  
methanol production

No. of trains: 1

Final product: C<sub>5</sub>+ 13,190 BPD  
LPG 2,218 "

i) Block Flow Diagram

See Fig. 8-2-7.

ii) Process Description

MTG Plant consists of the following process steps:

- . Dust Removal and 1st Compression
- . CO-Shift Conversion/COS Hydrolysis
- . Acid Gas Removal
- . 2nd Compression
- . Methanol Synthesis
- . MTG Synthesis
- . MTG Gasoline Separation
- . Alkylation

a) Dust Removal and 1st Compression

The raw gas leaving the gasifier at 4 kg/cm<sup>2</sup>G contains 50 kg/Nm<sup>3</sup> dust.

The dust in the raw gas is reduced to 5 mg/Nm<sup>3</sup> and this gas is compressed to 20 kg/cm<sup>2</sup>G.

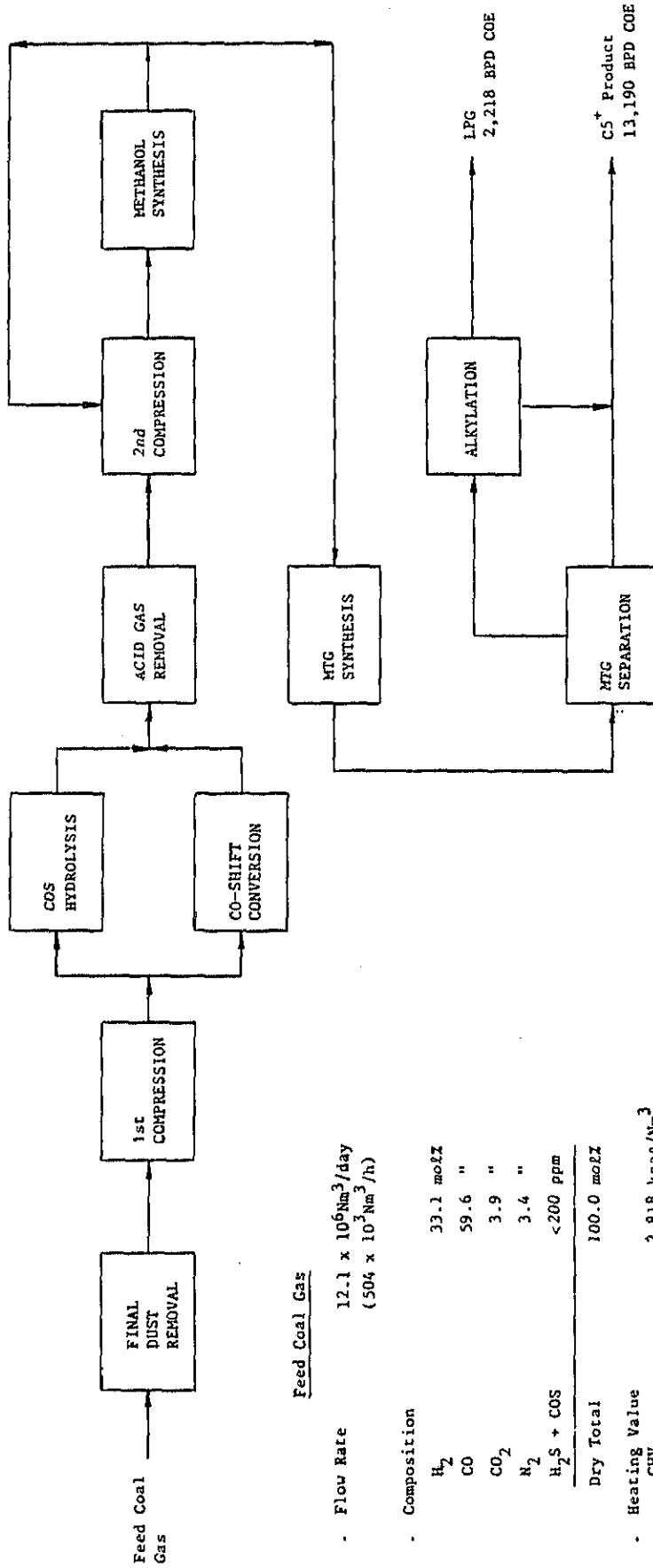
b) CO-Shift Conversion/COS Hydrolysis

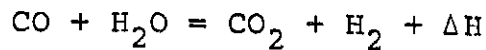
In order to adjust hydrogen/carbon monoxide ratio as required for the methanol production, 55% of the raw gas goes to CO-shift conversion where CO is decreased to 1% by the following reaction.

Fig. 8-2-7

BLOCK FLOW DIAGRAM

PRODUCTION OF GASOLINE FROM METHANOL





To keep the catalyst bed at a proper temperature, process steam or product gas is used to dissipate the reaction heat.

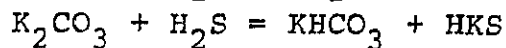
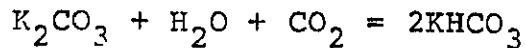
Shift conversion hydrolyzes much of COS to H<sub>2</sub>S.

The rest of raw gas goes to the COS hydrolysis unit where COS is also hydrolyzed into H<sub>2</sub>S.

Both gas streams are then mixed together after being cooled by-generating steam, and introduced to the acid gas removal unit.

c) Acid Gas Removal

Acid gases such as H<sub>2</sub>S and CO<sub>2</sub> are to be removed from the raw gas by hot potassium carbonate (HPC) solution.



The raw gas containing about 27% of CO<sub>2</sub> and 200 ppm of H<sub>2</sub>S is fed to the absorber where CO<sub>2</sub> and H<sub>2</sub>S gases are scrubbed by HPC solution, resulting in 3.5% of CO<sub>2</sub> and a few ppm of H<sub>2</sub>S. The treated raw gas leaving the top of the absorber goes to the 2nd compression while the rich HPC solution from the bottom is sent to a regenerator for acid gas stripping.

d) 2nd Compression

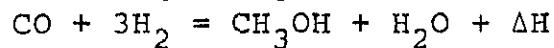
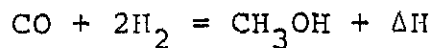
After acid gas removal, the treated raw gas is compressed to about 50 kg/cm<sup>2</sup>G.

e) Methanol Synthesis

The methanol synthesis loop comprises a recycle gas compressor, catalytic reactor, heat exchanger, condenser and separator.

The synthesis gas discharged from the compressor is fed to the catalytic reactor at about 230°C after heat exchange with the reactor effluent.

The reactor consists of multiple catalyst tubes in which the synthesis gas is converted to methanol by the following reactions:



As the catalyst cannot tolerate the adiabatic temperature rise of this exothermic reaction, the heat is dissipated by generating 40 kg/cm<sup>2</sup>G steam on the shell side of the reactor. The reactor effluent gas at 48 kg/cm<sup>2</sup>G and 260°C is cooled by heat exchange with recycle gas and cooling medium down to ambient temperature.

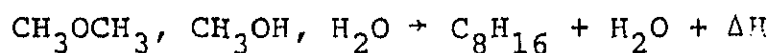
The produced methanol is separated from the unconverted gas in the separator and then sent to the distillation section via a depressuring valve. The unconverted gas is recycled and mixed with the raw gas as mentioned before. A small portion of the recycle gas is purged to limit the build-up of inerts in the loop. Purged gas is used as internal fuel.

f) MTG Synthesis

MTG Synthesis unit comprises two fixed-bed reactors (a DME reactor and a MTG reactor) and a heat exchanger, separator, recycle gas compressor. After pumped to about 30 kg/cm<sup>2</sup>G, the methanol is heated to about 300°C by heat exchange with the effluent from the MTG reactor. The evaporated methanol is then introduced to the DME reactor where conversion takes place and dimethylether is produced by the following reaction:



The effluent leaving the DME reactor at 400°C is combined with the recycle gas and fed to the MTG reactor at about 340°C. In the MTG reactor, the feed is converted to gasoline with a small amount of gaseous hydrocarbon and water by the following reaction:



As the exothermic reaction takes place in the reactor, cooled recycle gas is added to the MTG reactor inlet for temperature control. The effluent leaving the MTG reactor at 450°C is cooled by heat exchange with the feed to the DME reactor, and sent to the vapor-liquid separator via a steam generator and a cooler. The effluent cooled to ambient temperature is separated to hydrocarbon gas, gasoline and water.

A large part of the hydrocarbon gases from the separator is compressed and recycled to the reactor section, and the rest is sent to the stabilizer in the MTG gasoline separation unit.

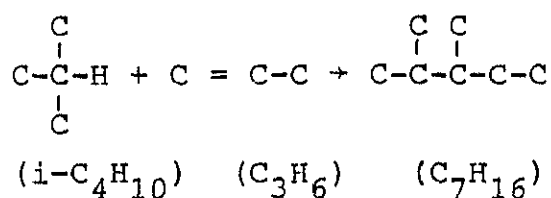
g) MTG Gasoline Separation

The MTG gasoline containing light hydrocarbons goes to the middle of the stabilizer after heat exchange with the stabilized gasoline from the stabilizer bottom.

The vapor from the stabilizer top is cooled by the overhead condenser, and propane and butane are condensed. These recovered LPG fractions are supplied to the alkylation unit. The stabilized gasoline is cooled down to ambient temperature and then stored in a tankage. The off gas from the overhead receiver is utilized as fuel gas outside the system.

h) Alkylation

LPG fractions recovered from the overhead receiver contains also propylene, butylene and isobutane which are important as blendstocks for a high octane alkylate gasoline. The following equation shows the formation of alkylation gasoline:



The raw LPG is mixed with recycled iso-butane, and introduced to the reactor where LPG is converted into butane alkylate (RON: 92) and propylene alkylate (RON: 97).

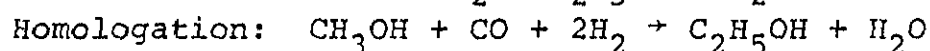
After separated from the propane LPG, alkylate gasoline is blended to MTG gasoline and sent to the tankage.

## (7) Others

The processes producing methanol derivatives such as ethanol, ethylene, acetic acid, ethylene glycol and MTBE are briefly discussed in this chapter. Among these processes, the acetic acid and MTBE processes have been commercialized while others are under development.

### 1) Ethanol Synthesis

There are two routes of ethanol synthesis; direct route and indirect one via methanol homologation.



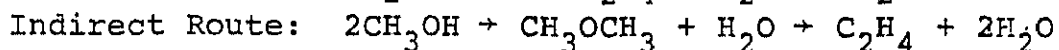
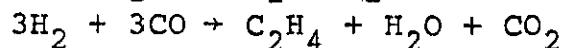
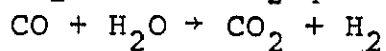
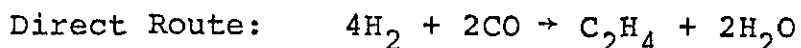
As for the direct route, Ruhrchemie and Kellogg presented in the 1950s Fe based catalysts of F/T type, while R&D on rhodium catalysts has recently been becoming popular.

As for the methanol homologation process, ethanol is synthesized from methanol and synthesis gas with ruthenium-cobalt or ruthenium-iron to which iodine or phosphorus is added.

The processes of this type were announced by such companies as Gulf, Exxon, UCC, Celanese and MGC. While these processes are designed to produce fuel ethanol or ethylene feedstocks, reductions in synthesis pressure and improvement of selectivity seem to form critical factors of their commercialization.

## 2) Ethylene Synthesis

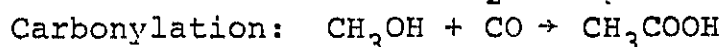
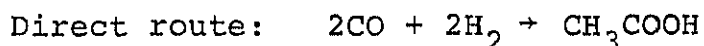
To synthesize ethylene from synthesis gas, there are direct and indirect processes.



With the use of zeolite catalysts developed by Mobil, and under conditions of 370°C and 2 atm, ethylene can easily be produced. MTG technology, which offers a foundation of the aforementioned cracking technology with similar catalysts, is now entering the stage of commercialization in New Zealand and expected to be the first commercialized process to synthesize ethylene from synthesis gas.

## 3) Acetic Acid Synthesis

Similar to ethanol synthesis, there are two routes of acetic acid synthesis. One is direct route and the other is indirect one via methanol carbonylation.



While a process to synthesize acetic acid via methanol was commercialized in 1960 by BASF, it could not gain popularity because it requires pressure as high as about 700 atm and because materials resistant to corrosive atmosphere under the presence of the acetic acid and carbon monoxide had not yet been developed at the time. In 1970, a low-pressure synthesis process developed by Monsanto which employs rhodium catalysts under the pressure of about 30 atm, is now surpassing the ethylene oxidation process which has conventionally served as the mainstream.

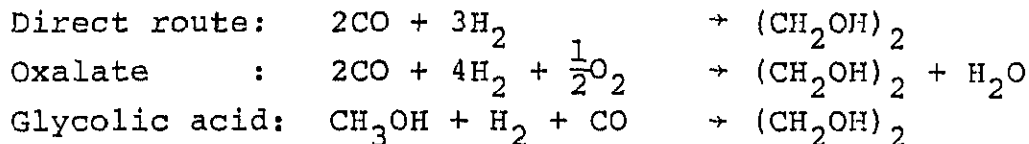
On the other hand, direct processes currently under development in such firms as UCC also employ rhodium catalysts. With this type of processes, synthesis reaction takes place under pressure as high as 300 atm associating too many side-reactions. Therefore, it is essential to develop catalysts featuring excellent selectivity and reduce

pressure. Using similar catalysts, it is possible to synthesize even high carboxylic acid, such as propionic acid and butyric acid.

#### 4) Ethylene Glycol Synthesis

There are three routes of ethylene glycol synthesis.

They are:



The glycolic acid process consists of three steps, including carbonylation, esterification and hydrogenation. The process had been in commercial operation from 1940 through 1967 by Du Pont, which employed a cobalt oxide catalyst and operating conditions of 500-700 atm and 200°C.

Recently Chevron has succeeded in low-temperature, low-pressure operations in the carbonylation step using HF instead of sulfuric acid, thus developing a superior process featuring a yield of 95% to Du Pont's. However, expensive HF and the need for additional facilities to offer separation, recovery and recycling resulted in diminishing economic feasibility of the new process.

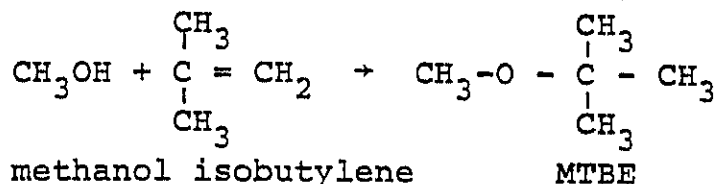
While ethylene glycol is currently produced with the ethylene oxide/glycol process, industrialization of synthesis gas routes (indirect route via methanol in the medium run and direct route in the long run) are expected as a result of R&D activities in the coming years.

#### 5) Methyl Tertiary Butyl Ether (MTBE)

There is considerable interest within the petroleum industry in the use of MTBE as a high-octane blending agent for use in motor gasoline. The material has been produced and marketed in Italy and the Federal Republic of Germany (West Germany) for several years, and recently a large plant has started production in the United States. Plans for a number of additional U.S. installations have been announced, and there are press reports of future projects.



MTBE is made by the following equation:



A few of the properties pertinent to its value in gasoline are compared with those of toluene as follows:

	<u>MTBE</u>	<u>Toluene</u>
Boiling point at 1 atm	55.2°C (131.4°F)	110.6°C (231.1°F)
Solubility in water		
at 20°C	4 wt%	0.05 wt%
Average blending octane number, $\frac{(R+M)}{2}$ , in lead-free gasoline pools	109	107

The octane performance of MTBE in gasoline blends depends on a number of factors relating to the specific situation in a given refinery. The blending octane number increases as the octane rating of the base stock decreases, as the saturated hydrocarbon content of the base stock increases, and as the MTBE concentration decreases.

## 8-3 Technology for Electric Power Generation

### (1) Utilization of Banko Coal for Conventional Coal Firing System

#### 1) Conventional Coal-Firing System

In the conventional application of a burner and fuel burning controlled, efficient conversion of the chemical energy of coal into heat energy which, in turn, is transferred to the heat-absorbing surfaces of the steam generator.

To do this, coal-firing systems introduce the coal and air for combustion, mix these reactants, ignite the combustible mixture and distribute the flame envelope and the products of combustion.

In the practical application of a burner and fuel burning system to a boiler, coal-firing system is categorized, as shown Fig. 8-3-1.

#### i) Horizontally fired system

In horizontally fired systems the coal is mixed with combustion air in individual burner registers. The coal and primary air are introduced tangentially to the coal nozzle, thus imparting strong rotation within the nozzle. Adjustable inlet vanes impart a rotation to the preheated secondary air from the windbox. The burners are located in rows, either on the front wall only or on both front and rear walls.

(Fig. 8-3-2)

#### ii) Tangentially fired systems

The tangentially fired system is based on the concept of a single flame envelope. Both fuel and combustion air are projected from the corners of the furnace along a line tangent to a small circle, lying in a horizontal plane, at the center of the furnace. Intensive mixing occurs where these streams meet. A rotative motion, similar to that of a cyclone, is imparted to the flame body, which spreads out and

fills the furnace areas. (Fig. 8-3-3)

iii) Vertically fired systems

Vertically fired systems are used today only to fire solid fuels that are difficult to ignite, such as coals with moisture-and-ash-free volatile matter less than 13 percent. The firing concept is the vertical arrangement of the burners in the arches.

(Fig. 8-3-4)

iv) Spreader stokers for coal firing

Spreader stokers use the combined principles of pulverized-coal and stoker firing in that fines are burned in suspension and the larger particles are burned on the grate. Feeding and distributing mechanisms continually project coal into the furnace above an ignited fuel bed. (Fig. 8-3-5)

Fig. 8-3-1

CONVENTIONAL COAL-FIRING SYSTEM

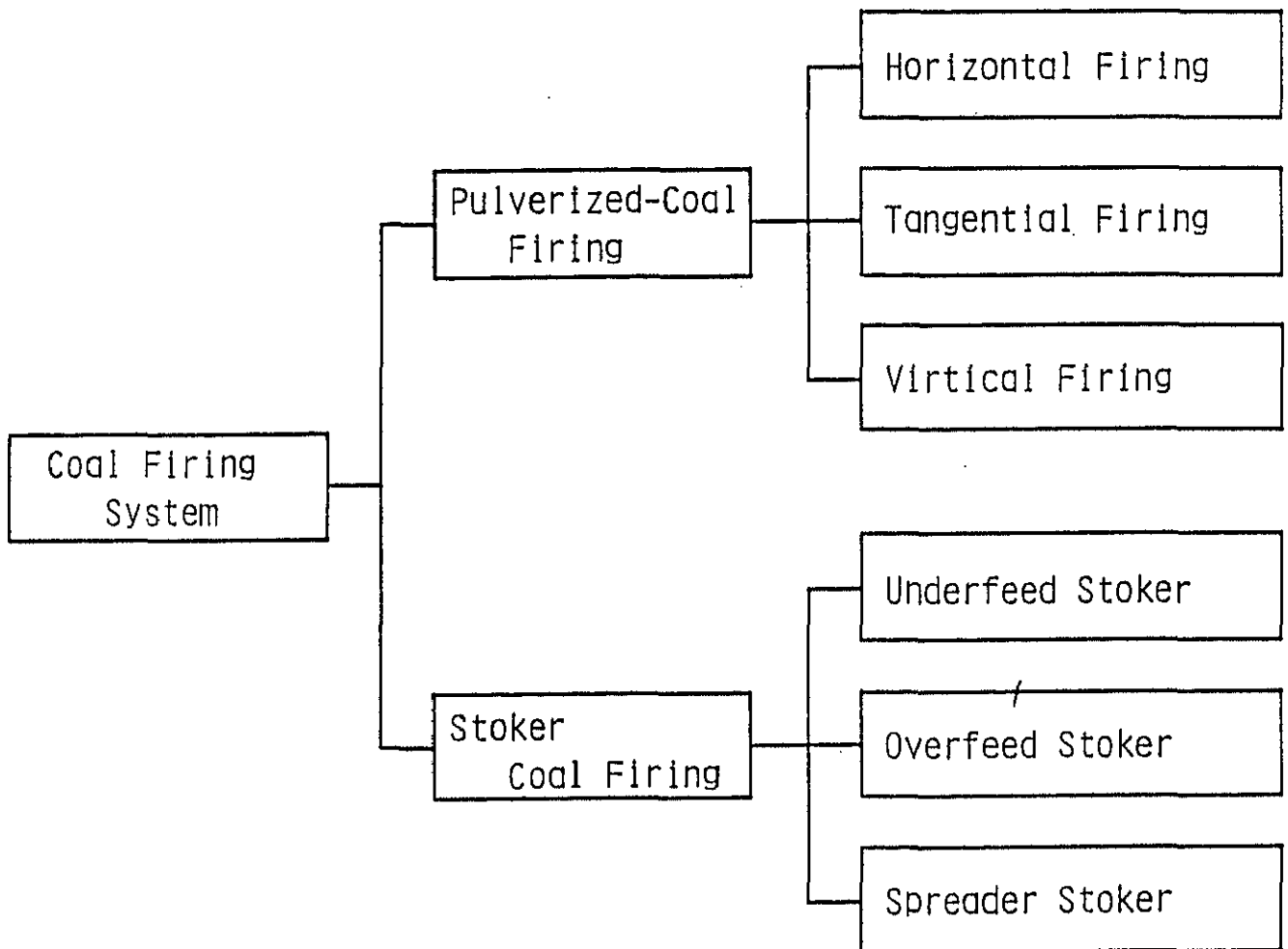


Fig. 8-3-2 HORIZONTALLY FIRED SYSTEM  
(SIDE VIEW)

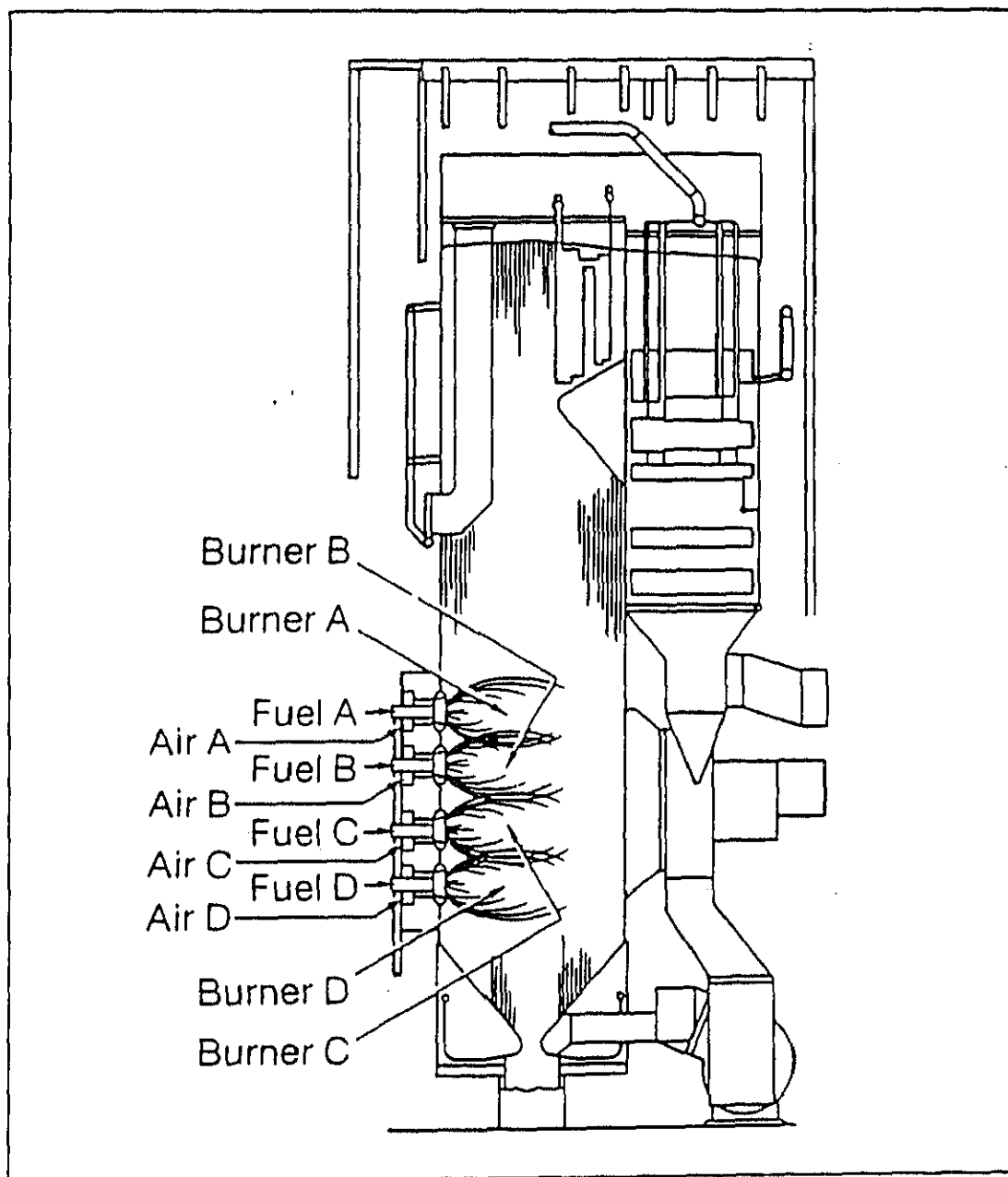


Fig. 8-3-3 . TANGENTIALLY FIRED SYSTEM  
(PLAN VIEW)

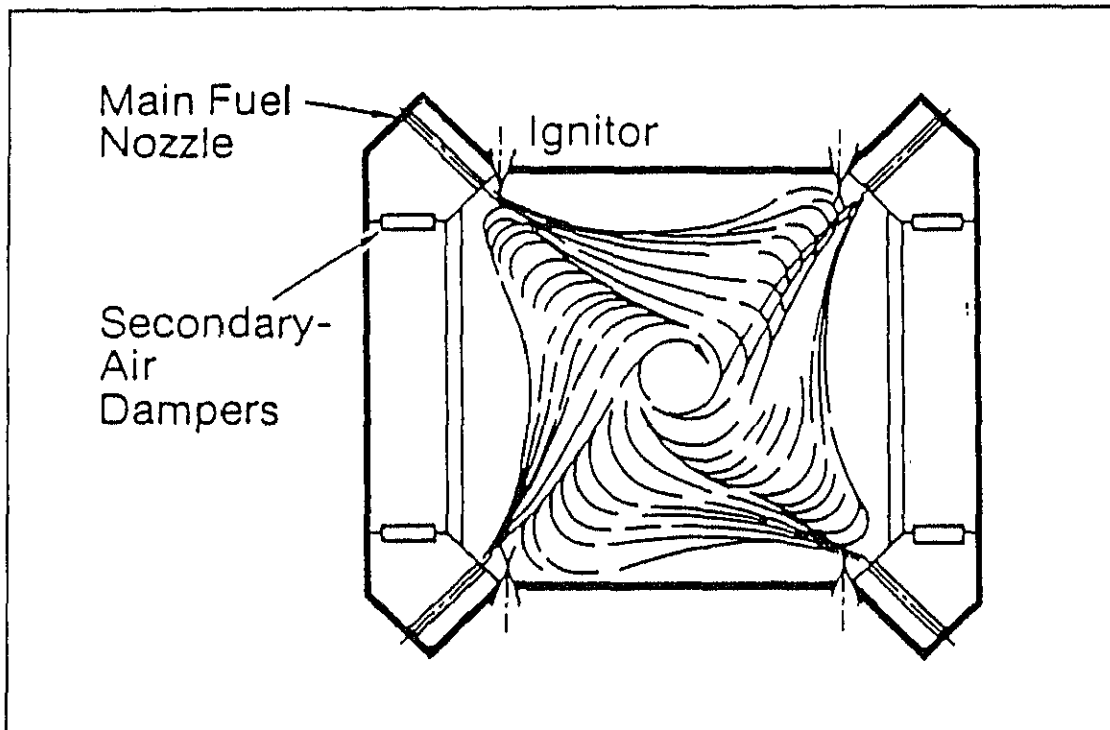


Fig. 8-3-4 VERTICALLY FIRED SYSTEM  
(SIDE VIEW)

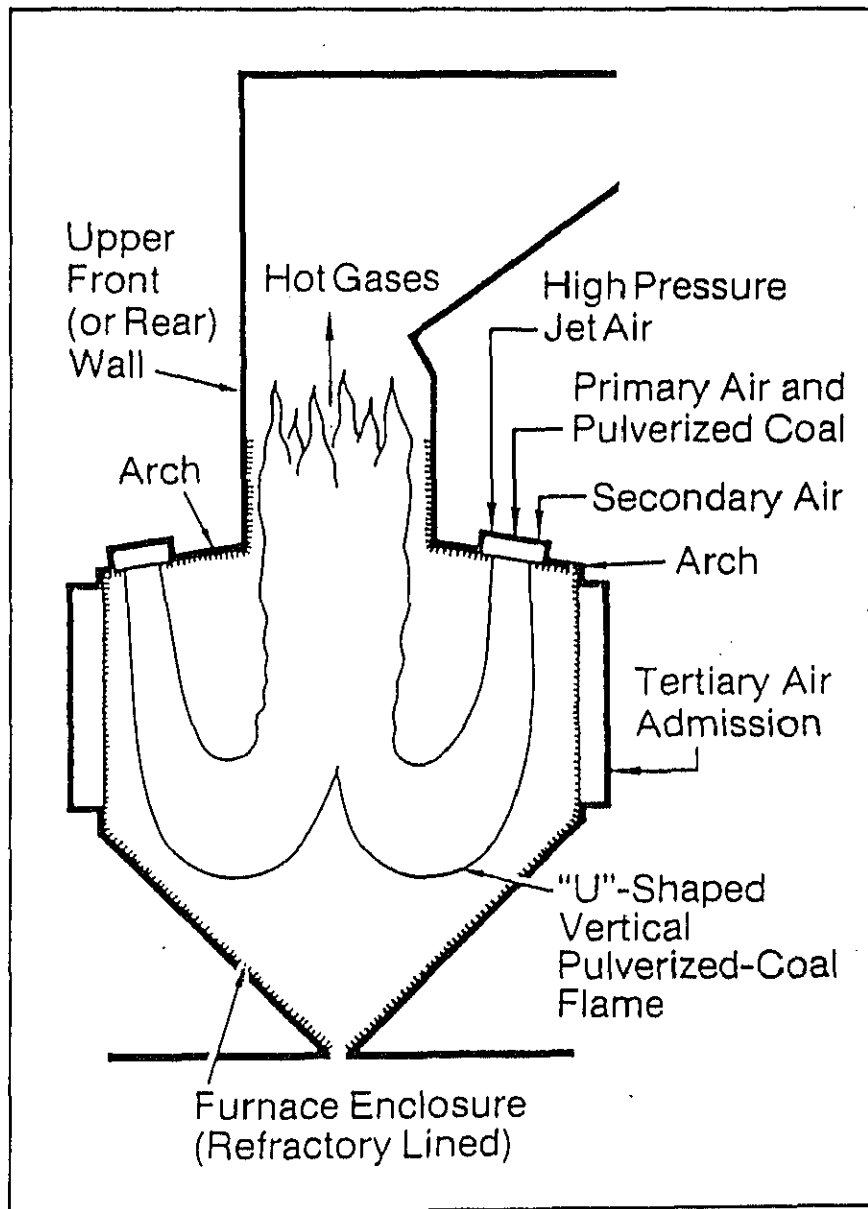
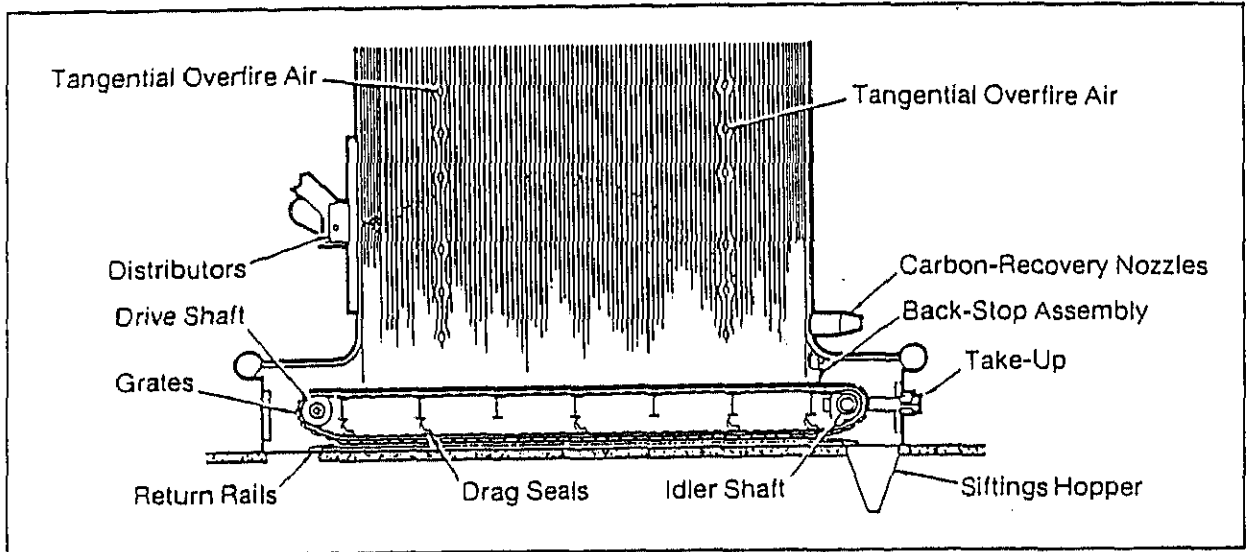


Fig. 8-3-5

SPREADER STOKER,  
CONTINUOUS ASH DISCHARGE GRATE



2) Possibility of Banko coal utilization by conventional coal firing system

i) Combustion characteristics of Banko coal

The Banko coal is likely to exhibit similar combustion characteristics to some of the Australian and U.S.A. brown coals currently used in power station boilers. However, the following indicative analysis of the Banko coal, particularly the high sodium levels found in the lower seams, suggests that it should be regarded as a "problem" coal which could result in severe fouling and slagging in utility boilers.



Table 8-3-1 Ash Analysis of Indonesian Coals

	<u>Air Laya</u>	<u>Banko</u>	(% wt)
SiO <sub>2</sub>	59.4	20.0	- 61.0
Al <sub>2</sub> O <sub>3</sub>	24.7	19.0	- 41.0
Fe <sub>2</sub> O <sub>3</sub>	4.6	0.8	- 17.0
TiO <sub>2</sub>	0.8	0.8	- 17.0
CaO	3.1	0.6	- 1.3
MgO	1.7	0.2	- 6.6
Na <sub>2</sub> O	2.5	0.2	- 20.0
K <sub>2</sub> O	0.5	0.2	- 0.6
Mn <sub>3</sub> O <sub>4</sub>	-	0.04	- 0.79
P <sub>2</sub> O <sub>5</sub>	0.4	0.04	- 0.79
SO <sub>3</sub>	2.3	0.58	- 13.0

The problem is essentially caused by the detrimental ash characteristics which cause slagging and fouling to occur within the boiler. In this context, "slagging" refers to the deposition of liquid or molten ash on tubes within the radiant section of the boiler and is a function of the melting properties of the ash. The other major form of ash deposition on heat transfer surfaces within the boiler is termed "fouling". Fouling usually occurs in the convective superheater and reheater tube banks and is instigated by the condensation of alkali salts (Na<sub>2</sub>SO<sub>4</sub>, CaSO<sub>4</sub>) which are volatilised in the flame.

- ii) Combustion technology for coal of high sodium-in-ash. In the U.S.A., firing of low grade coals on pulverised fuel units got off to a slow start and it was the 1960s before the first units were being put into operation. These were designed with sodium in ash contents of up to 5% in mind and major operational problems were experienced when higher percentage sodium in ash fuels were actually encountered. The fuels to be used for the series of new generation

boilers include coals of up to 10% sodium in ash. To cater for these difficult coals much more conservative boiler designs are being used with lowly rated furnaces and specific provisions for flexibility in furnace operation. It is also important to note that the major deposits currently being considered in the West do not have sodium in ash contents as high as 10%. This level can therefore be considered as the practical limit for plant design in the U.S.A., since it is unlikely that experience will be gained with worse fuels having higher sodium in ash contents, as major deposits of such fuel have not been discovered to date.

The question of slagging and fouling with this type of coal can be summarised in the following way:

- a) In the light of world experience it can be said that proven technology and considerable operational experience are available for the use of fuels with up to 5% sodium-in-ash.
- b) Technology has been developed but experience is still being gained with fuels containing 10% sodium-in-ash.
- c) Fuels containing more than 10% sodium-in-ash are now being considered for new power projects but it will be the late 1980s before they become commercial propositions.

The potential problems are greater if this type of coal is destined for existing power stations which have been designed for coals with better combustion characteristics e.g. Java power station (Air Laya coal). In such a case the severe fouling characteristics could reduce station output, increase maintenance costs and reduce plant availability dramatically unless considerably modifications were carried out to the existing boilers.

iii) Coal utilization by fluidized-bed combustion

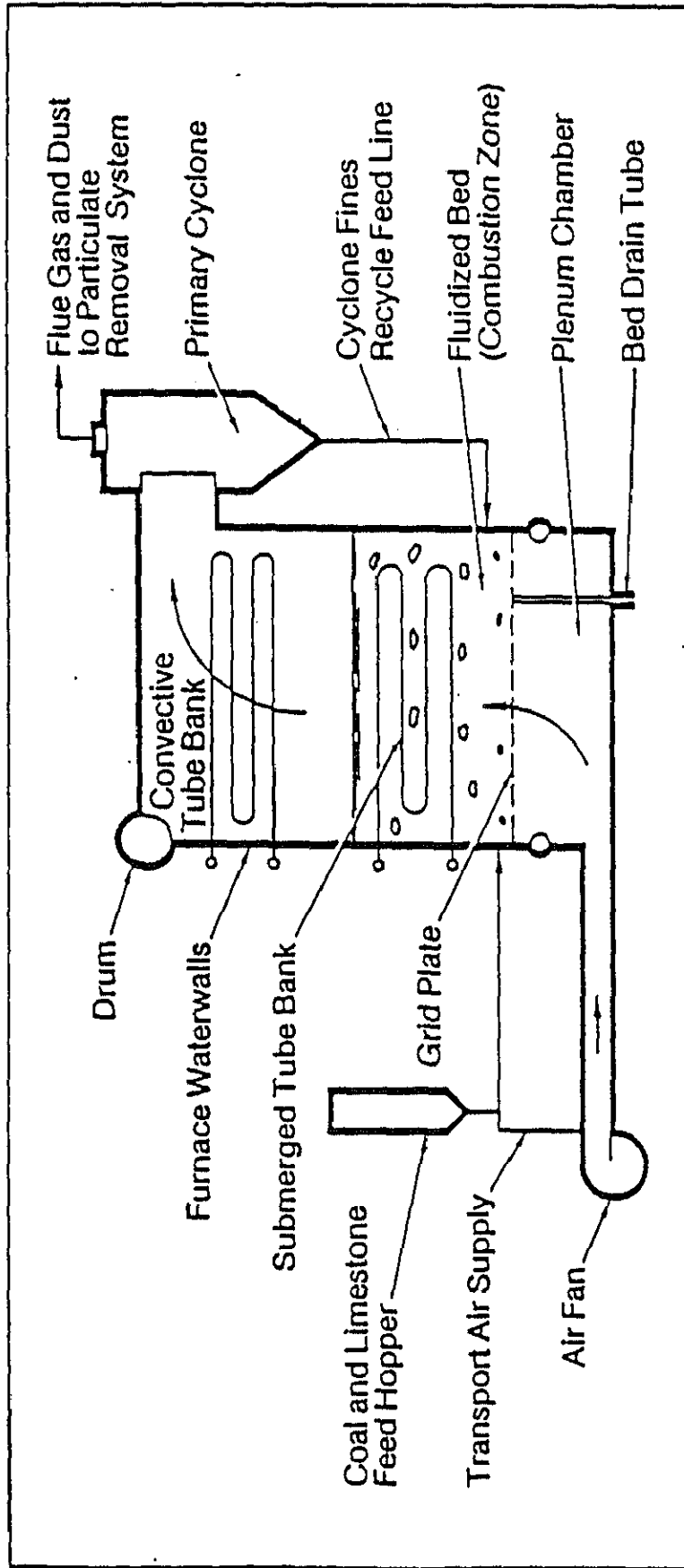
Though it is expected that modified technology of conventional coal firing system, available for the use of more than 10% sodium-in-ash, will be developed, fluidized-bed combustion system for electric power generation, which are under development, seem to be more prospective because the technology will be available for coal of high sodium-in-ash.

Fluidized-bed combustion can burn coal efficiently at a temperature low enough, in range of 700 - 900°C, to avoid many of the problems of conventional combustion. The outstanding advantage of fluidized-bed combustion (FBC), then, is its ability to burn high-sulfur coal in an environmentally acceptable manner without the use of flue-gas scrubbers. A secondary benefit is the formation of lower levels of nitrogen oxides compared to other combustion methods. In addition, low bed temperatures help eliminate the potential for slag formation on the water-cooled walls of the furnace (an operational advantage) and high heat-transfer in the fluidized bed permits a more economical boiler design. (Fig. 8-3-6)

Fluidized-bed combustion has been under development since the 1950s with the principal intent to demonstrate the economy of the coal/limestone system for sulfur-oxide control as compared with post-boiler wet scrubbing.

Fig. 8-3-6

SCHEMATIC DIAGRAM OF  
FLUIDIZED-BED-COMBUSTION BOILER



(2) Utilization of Banko Coal for Coal Gasification  
Combined-Cycle System

1) The Present situation of combined-cycle power generation

The so-called "combined-cycle power generation" consists in a combination of gas and steam turbines in order to improve the thermal efficiency and various systems of it have been proposed and tried with the development of gas turbine. (Fig. 8-3-7)

There is various combined-cycle power plants now in operation, using natural gas or oil as fuel as shown in Table 8-3-2, resulting in 20,000 MW of the total output, which is equally distributed in USA and Europe by respective 10,000 MW. In USA, the greater part (85%) of it is covered by the exhaust gas heat recovery cycle system, whereas in Europe the predominant part (94%) by the exhaust gas fired cycle system.

In Japan, four of large scale power plant, using LNG as fuel, covered by the exhaust gas heat recovery cycle are under construction as shown on Table 8-3-3. Since the appearance of conventional gas turbine of around 1,100°C as its inlet temperature, combined-cycle system, of which output is mainly composed of that of the gas turbine, has been investigated in order to utilize the characteristics of this high-efficiency gas turbine, so that it is now possible to expect a more remarkable improvement in thermal efficiency by means of a system, in which steam is made to be generated with heat recovered from the gas turbine exhaust gas which is so-called exhaust gas heat recovery cycle.

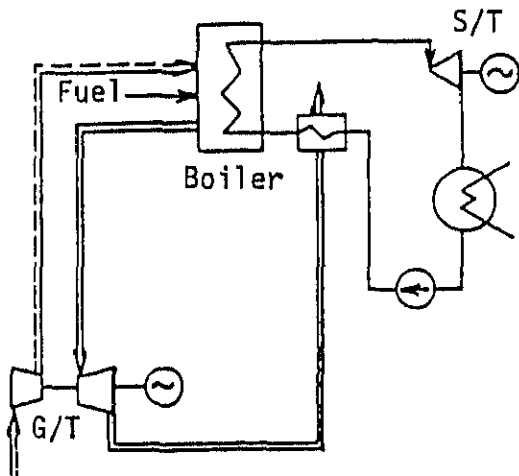
2) Outline of features of coal gasification combined-cycle power plant

On Coal Gasification Combined-Cycle (CGCC) for power generation, investigation and technical research has been world-widely promoted.

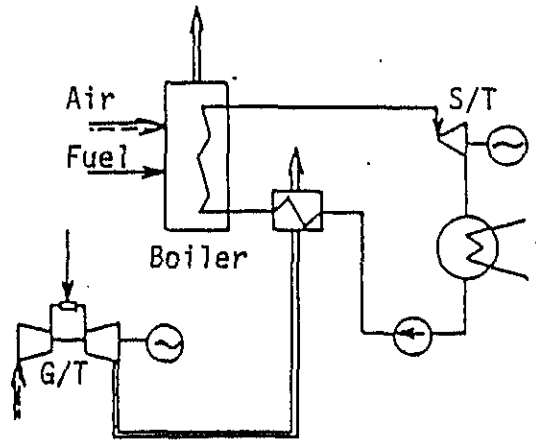
The technology of CGCC has been developed as the CGCC is environmentally-acceptable and more advanced power generation system, and therefore the CGCC is required to have reliability and economic performance for its application to

Fig. 8-3-7 TYPE OF COMBINED-CYCLE PLANT

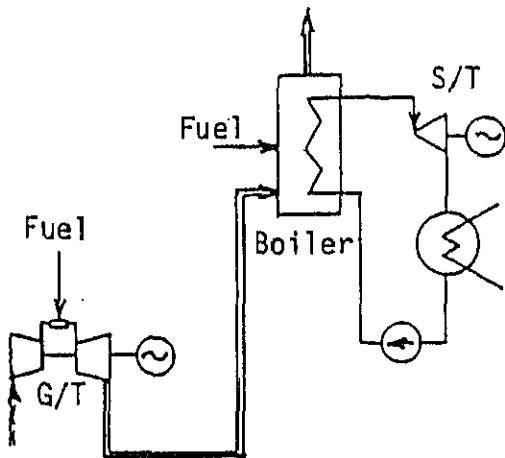
SUPERCHARGED BOILER CYCLE



SIMPLE FEED WATER HEATING CYCLE



EXHAUST GAS FIRING CYCLE



EXHAUST GAS HEAT RECOVERY CYCLE

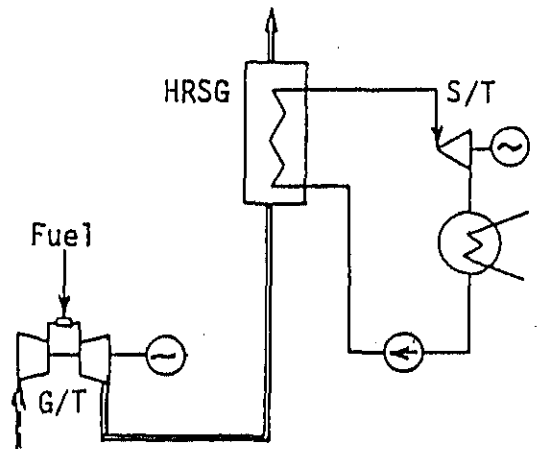


Table 8-3-2 WORLD-WIDE COMBINED-CYCLE POWER PLANT INSTALLATIONS

District	Plant type	No. of units	Total output (MW)	Commencement of operation
U.S.A.	Feedwater heating	2	70.0	1949
	Exhaust gas fired	10	1392.5	1958
	Exhaust heat recovery	35	7990.6	1968
	Subtotal	47	9453.1	-
Europe incl. U.S.S.R.	Supercharged boiler	5	941.0	1963
	Feedwater heating	1	312.0	1964
	Exhaust gas fired	26	8862.0	1964
	Exhaust heat recovery	4	265.5	1960
	Subtotal	36	10380.5	-
Asia	Exhaust gas fired	2	258.0	1968
	Exhaust heat recovery	2	640.0	1979
	Subtotal	4	898.0	-
Total		87	20731.6	-

Table 8-3-3 COMBINED-CYCLE POWER PLANT IN JAPAN

<Exhaust Gas Heat Recovery Cycle>

CLIENT	PLANT OUTPUT	FUEL	COMMISSIONING
JAPAN NATIONAL RAILROAD CO. KAWASAKI P/S	140 MW	DISTILLATE OIL	1981.3
NIPPON STEEL CO. KAMAISHI WORKS	25 MW	BFG, COG	1982.2
TOHOKU ELECTRIC POWER CO. HIGASHI NIIGATA P/S	1,090 MW	LNG	1984.12/1985.10
TOKYO ELECTRIC POWER CO. FUTTSU P/S	1,000 MW 1,000 MW	LNG LNG	1985.9 1985.12
CHUBU ELECTRIC POWER CO. YOKKAICHI P/S	560 MW	LNG	1986.5



power generation.

In an integrated electric power-plant application, the gasification system is part of a two-stage coal combustion process. In the first, or gasification stage, the coal is partially reacted with a deficiency of oxygen to produce a low-calorific-value (LCV) fuel gas that can be readily cleaned.

Pollution control before combustion is accomplished by partial oxidation of the coal to produce a clean gas. Sulfur and ash are removed in the processes, minimizing problems caused by  $\text{Na}_2\text{O}$  in ash.

In the second stage, the cleaned fuel gas is burned in a boiler and/or gas turbine for the generation of electric power. Heat produced in the gasification stage is recovered by generating steam; therefore, the gasifier has a common need with the power boiler for heated air and feedwater. (Fig. 8-3-8, Fig. 8-3-9)

The topic of technical development of the CGCC is on technology of advanced gas turbine as well as gasification and dry gas-cleanup system.

For gas turbine inlet temperature, thermal efficiency of the CGCC is promised to be improved as the temperature is rising, as shown in Fig. 8-3-10.

Table 8-3-4 shows typical coal consumption, comparing between CGCC and coal-fired steam cycle.

### (3) Development of Coal Gasification Combined-Cycle Power Plant

The Coal Gasification Combined-Cycle which has advantage of performance and environmental effect has been worldwidely developed to raise security of power source.

In USA, Cool Water Project of CGCC is promoted and plant operation is started in this year. (Fig. 8-3-11)

The net thermal efficiency of Cool Water Project is estimated to be approximately 31 percent, comparatively lower

Fig. 8-3-8

### INTEGRATED COAL-GASIFICATION/COMBINED - CYCLE POWER PLANT

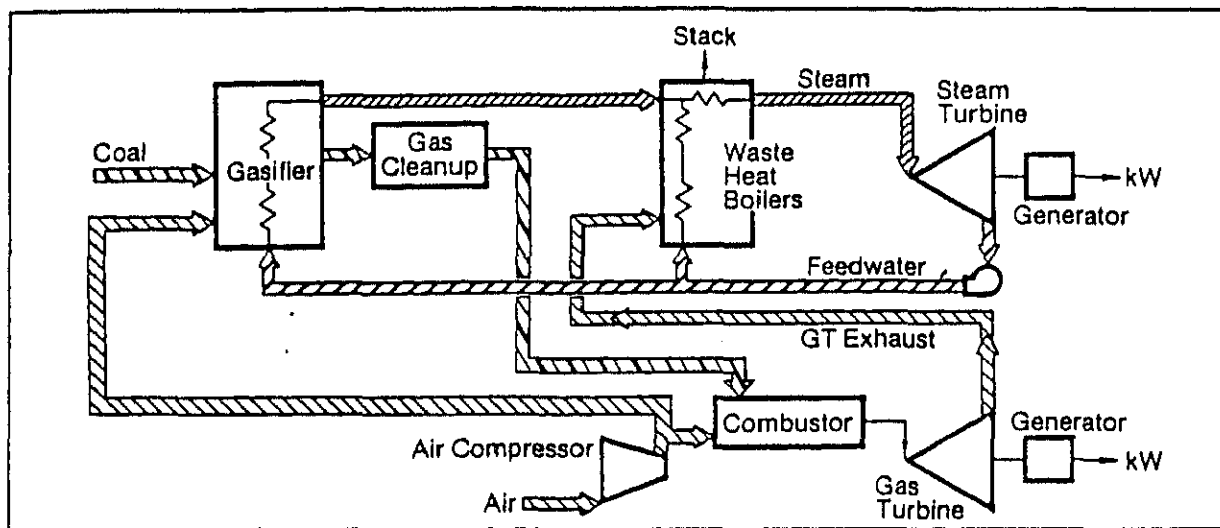


Fig. 8-3-9

### INTEGRATED COAL-GASIFICATION/STEAM-CYCLE POWER PLANT

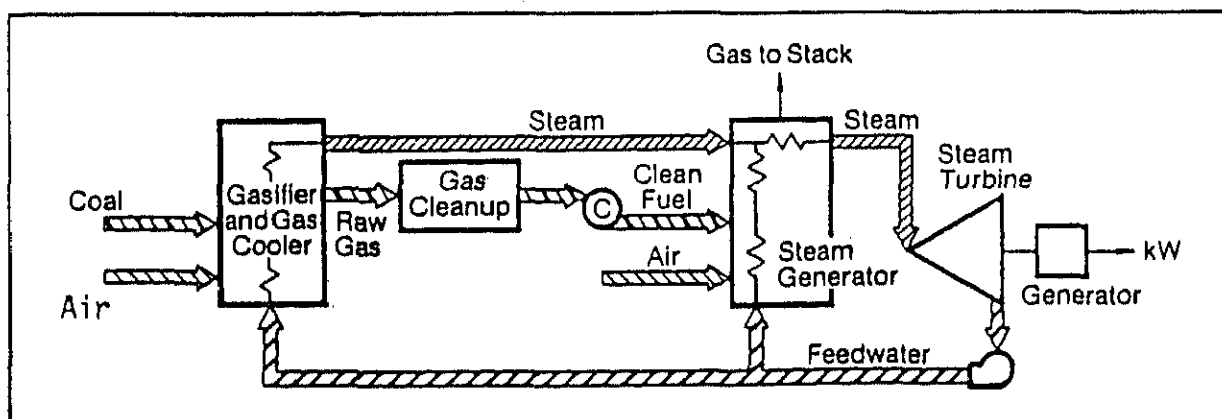


Fig. 8-3-10 THERMAL EFFICIENCY OF  
COAL GACIFICATION COMBINED-CYCLE

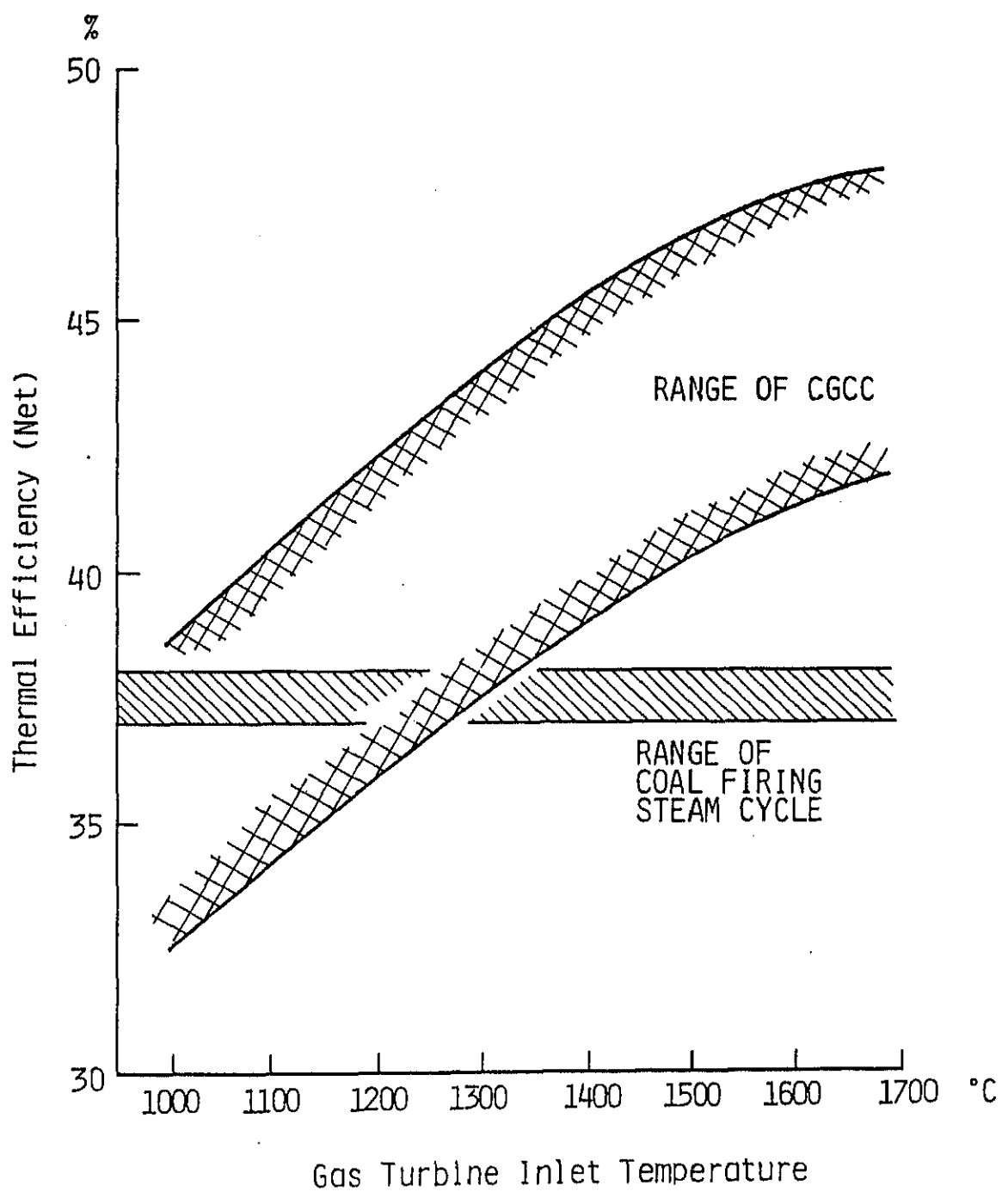


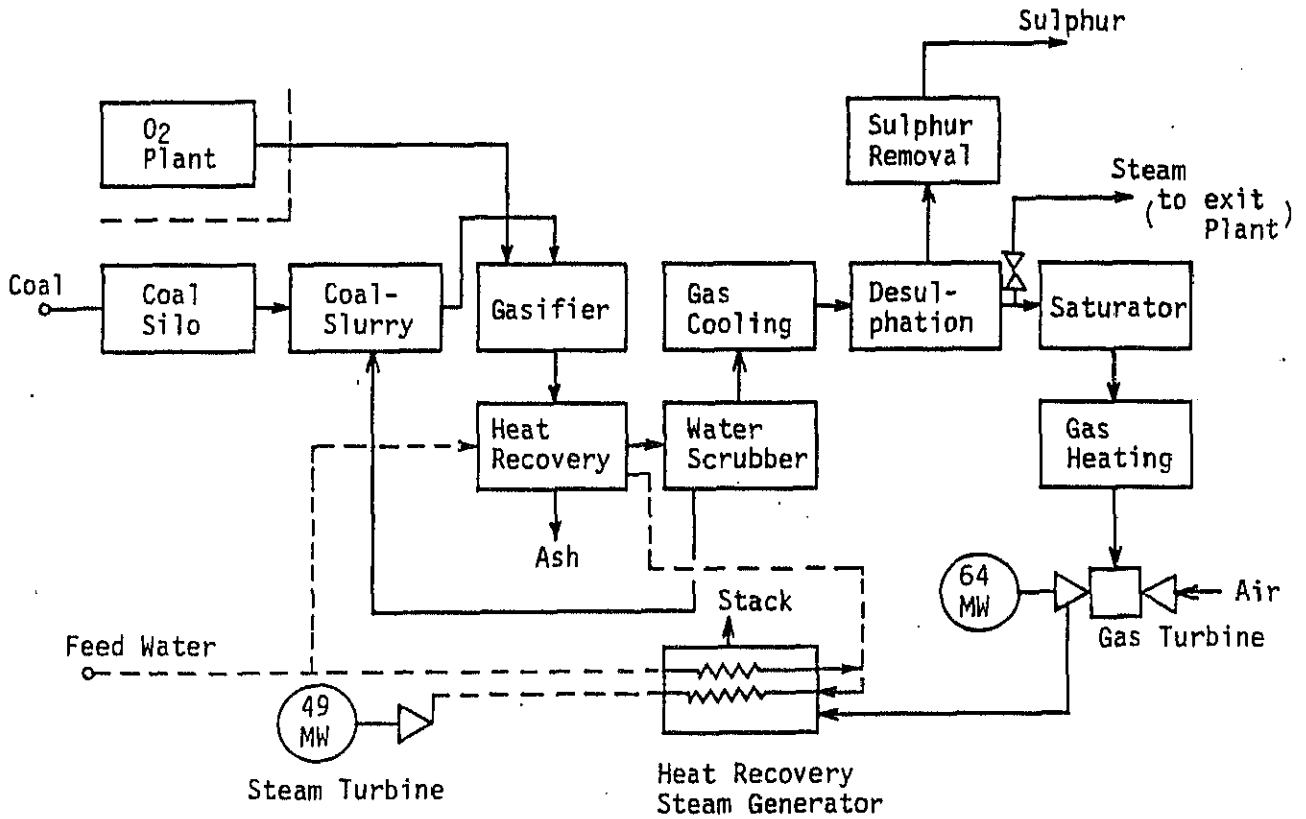
Table 8-3-4 THERMAL EFFICIENCY AND ANNUAL COAL CONSUMPTION

		Coal Gasification Combined-Cycle (CGCC) Power Plant	Coal-Fired Steam Cycle Power Plant
Thermal Efficiency (Net)	%	43 (1,300°C G/T)	37
Heat Rate	kcal/kWh	2,000	2,324
Annual Heat Consumption	kcal/kW·Y	$12.26 \times 10^6$	$14.25 \times 10^6$
Annual Heat Consumption (1,000 MW)	kcal/Y	$12.26 \times 10^{12}$	$14.25 \times 10^{12}$
Annual Coal Consumption (1,000 MW)	t/Y	$1,890 \times 10^3$ $\ominus 300 \times 10^3$	$2,190 \times 10^3$ (base)

(Note) Annual Capacity Factor = 70 %  
Coal HHV = 6,500 kcal/kg

Fig. 8-3-11

OUTLINE OF  
COOL WATER CGCC PROJECT



Site	Cool Water P/S (SCE)
Coal Feed	1,000 t/d
Power Output	113 MW
Gas Turbine	64 MW
Steam Turbine	49 MW

than 37 percent of conventional coal-firing steam cycle system, because the project applies coal water slurry system, wet gas-cleanup system and oxygen as oxidant. However the project is notably watched through the world because it is the first demonstration plant of CGCC system.

In Japan, CGCC development and technical research is promoted with 40 t/d fluidized-bed gasification test plant by Sun Shine Project and 2 t/d entrained-bed gasification test plant by CRIEPI (Central Research Institute Electric Power Industry). (Table 8-3-5)

The development of Advanced Gas Turbine of 1,300°C class and dry clean-up system has been promoted in Japan as well as in USA, one of which is by Sunshine Project of Japan.

#### (4) Prospects of Coal Gasification Combined-Cycle Power Plant

Outline of expected Coal Gasification Combined-Cycle Power Plant is described in Fig. 8-3-12 and Table 8-3-6.

In these rough sketch, the design condition is set that gas turbine is 1,300°C class, gas clean-up is dry system and steam cycle is integrated-system between gasification plant, gas clean-up and heat recovery steam generator of gas turbine.

Rough performance data is shown for two case of 300 MW class and 1,000 MW class.

The technical possibility of CGCC system for Banko coal utilization depends on availability of CGCC system for high sodium-in-ash and a wide range of ash fusion temperature as well as technology development for the design conditions set above.

According to the results of chip sample analysis and Shell data, sodium-in-ash is 12 percent as average, ranging from 0 to 40 percent, and ash fusion temperature is scattered from 1,150°C to 1,500°C or higher (not confirmed) by area and

Table 8-3-5 . CGCC DEVELOPMENT IN JAPAN

<40 t/d Fluidized-bed Coal Gasification>

Coal Feed	40 t/d
Gasifier	Pressure
	Temperature
Test Equipment	Gas Cleanup Test G/T. Combustion Test ..

<2 t/d Entrained-bed Coal Gasification>

Coal Feed	2 t/d
Gasifier	Pressure
	Temperature
Test Equipment	{ Gas Cleanup Test G/T Combustion Test }

Fig. 8-3-12 SCHEME OF COAL GASIFICATION  
 COMBINED-CYCLE POWER PLANT  
 (Integrated)

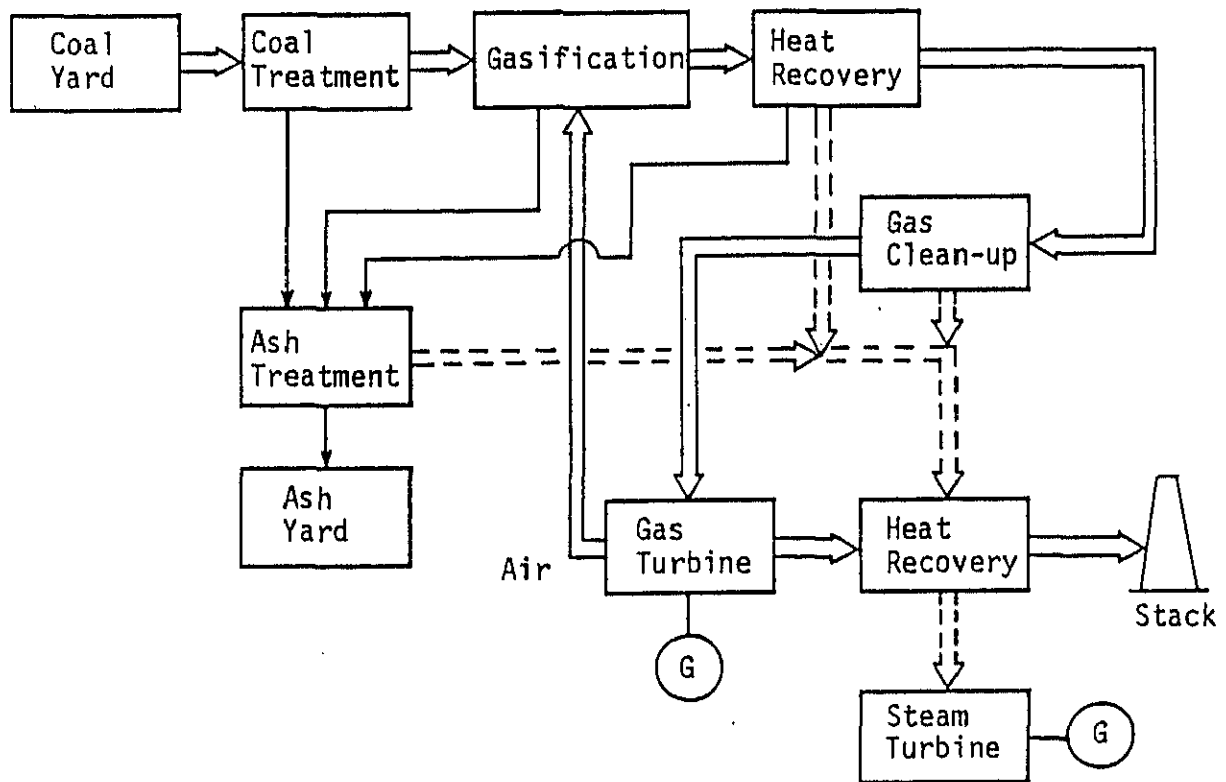




Table 8-3-6 OUTLINE OF EXPECTED COAL GASIFICATION  
COMBINED-CYCLE POWER PLANT

	300 MW Class	1,000 MW Class
Ambient Temperature Barometric Pressure	15 °C 760 mmHg	
Gas Turbine Clean-up System	1,300 °C Class Dry Clean-up	
Coal Feed (HHV)	82 t/h 6,300 kcal/kg	328 t/h 6,300 kcal/kg
Output		
Gas Turbine	138.5 MW x 1	138.5 MW x 4
Steam Turbine	140.5 MW x 1	283 MW x 2
Total	279 MW	1,120 MW
Circulating Water Flow	28,800 m <sup>3</sup> /h	115,000 m <sup>3</sup> /h

seam of sampling spot. The results of survey on coal gasification technology and features of the fluidized-bed combustion boiler suggest that the fluidized bed gasifier operating at around 1,000°C will be available for Banko coal, because low bed temperature held eliminates the potential for slag formation on the bed of gasifier and the water cooled surfaces of the heat recovery system. As conclusion, it is understood that the prospects of CGCC system for Banko depends on technical development of gas turbine and dry gas-cleanup system.

## 8-4 Technology for Methanol Engine

### (1) Fuel for Internal Combustion Engines

The internal combustion engines presently in use include three types; the gasoline engine (Otto cycle), diesel engine (Sabathe cycle), and gas turbine engine (Brayton cycle).

Fig. 8-4-1 illustrate features of each engine.

The Otto cycle, generally known as a gasoline engine, uses spark ignition. To prevent knocking, the engine uses fuel with high octane number. The Sabathe cycle known as a high-speed diesel engine uses compression ignition. Fuel with high cetane number is used for the diesel engine. Also, Brayton cycle is used for the gas turbine. Gas turbine features continuous combustion under a constant pressure and is knock-free.

Each of these engines is designed to heat and expand air through combustion of fuel and convert the expanding energy into mechanical work. Internal combustion engines feature as follows:

a) Thermal efficiency--Diesel (40%) > Gasoline (25%) > Gas turbine (15%)

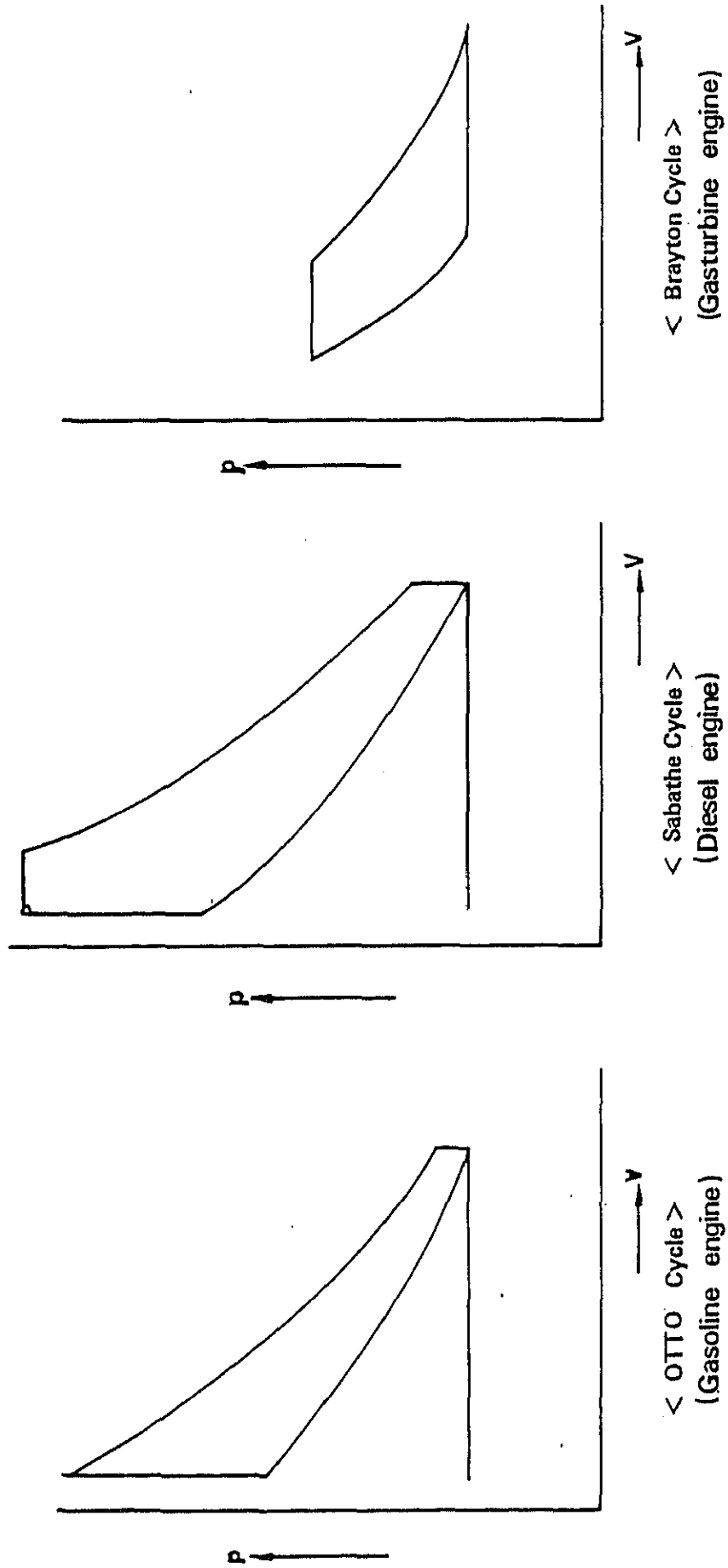
However, the gas turbines equipped with a heat recovery system posses thermal efficiency equal to diesel engines.

b) Weight/output--Diesel (4 kg/ps) > Gasoline (2 kg/ps) > Gas turbine (1 kg/ps)

The above explanation indicates reasons why the engine manufacturing industry sets targets on higher thermal efficiency through use of ceramics, a turbocharger and weight reduction of engine itself. Progress in improvement of each of the engines has been made at a speed unparalleled in the past, yet the relative efficiencies of each particular engine have not changed.

Consequently, compact automobiles and ships use a gasoline engine because of lightweight compact design and relatively high thermal efficiency. Large-sized vehicles (buses and trucks), large-sized ships and small scale of

Fig. 8-4-1 Theoretical P-V Diagram of Typical Internal Combustion Engines



fixed type power generators use a diesel engine with high thermal efficiency. Large-sized power generators and aircraft use a gas turbine which features high durability, low vibration, and light-weight and compact design.

The fuel used for these internal combustion engines is generally petroleum products such as gasoline, diesel oil, kerosene and heavy oil. Each type of engines is designed to attain the highest performance for each petroleum product, the gasoline engine for gasoline, diesel engine for diesel oil and gas turbine for kerosene and diesel oil.

Let's examine the role of fuel that plays in those internal combustion engines. As principle all combustible material can be used as fuel for internal combustion engines because fuel is simply required to heat and expand air. The gas turbine allows use of heavy oil, gas and even coal powders because it has a structure least affected by fuel. Considering the simple structure of the gas turbine, methanol is a fuel highly suited to this type of engine, (Fuel is injected into air compressed to a high temperature and pressure, and this mixture is ignited by spark. It is then guided into the turbine for expansion.).

The properties required of a fuel for internal combustion engines is, simply put, that it shall be combustible as mentioned above. In more concrete terms, however, considering the practical requirements of engines such as lightweight and compact design, high efficiency, reliability and durability, the fuel should have:

- a) Adaptability to a wide range of combustion characteristics,
- b) Appropriate calorific value per unit of fuel,
- c) Good lubrication characteristics,
- d) A low freezing point for use in cold areas,
- e) Anti-corrosion characteristics for protection of engine components.

Item c, d and e are the important factors in determining reliability and durability.

(2) Comparison of Methanol and Petroleum as Fuel for Internal Combustion Engines

Table 8-4-1 shows a comparison of fuel properties of methanol, gasoline and diesel oil.

As is clearly indicated in the table and from production processes, methanol has properties quite different from gasoline, and diesel oil. This stems from the difference in molecular structure and chemical composition between petroleum and methanol.

Show below are the subjects that must be studied to enable use of methanol in ordinary internal combustion engines.

- i) Calorific value per weight of methanol is 50% lower than gasoline and diesel oil.

To obtain the same output from a given engine which is designed for gasoline or diesel oil in use, the quantity of injection has to be double. Accordingly, some devices for better mix with air are required. This requires modification of injection and combustion systems, and larger injection devices and fuel tanks.

- ii) Because methanol has a relatively high octane number, it is possible to take a higher compression ratio in a spark ignition engine for higher combustion efficiency. The other hand, it tends to produce HCOOH in combustion chambers because of relatively low combustion temperature. To cope with this, it is necessary to improve acid resistance of materials of the combustion system.

- iii) Because of a low cetane number, ordinary compression ignition system can not be applied. Consequently, when methanol is used as fuel for a diesel engine, the following measures are recommended;

- a) To charge methanol and diesel oil at the same time into combustion chamber.
- b) To install a spark plug for assist of ignition.

Table 8-4-1 Characteristics Comparison of Methanol with Petroleum Fuel as Fuel of Internal Combustion Engines

ITEM	METHANOL	GASOLINE	LIGHT OIL	ETHANOL	JET FUEL	REFERENCE
Chemical formula	CH <sub>3</sub> OH	C <sub>5</sub> H <sub>12</sub> ~ C <sub>9</sub> H <sub>20</sub>	—	C <sub>2</sub> H <sub>5</sub> OH	—	
Molecular weight	32	(114)	—	45	—	
Specific gravity (20°C)	0.76	0.70	0.83	0.78	0.75	
Theoretical air/fuel	6.4	15.1	14.5	9.0	—	Theoretical air-fuel ratio to get perfect combustion
Heating value (Kcal/kg)	4,800	10,600	10,320	6,400	10,300	Lower value
Octane number	110	100	—	105	—	Research octane number higher number lower knocking
Cetane number	3	12	45 ~ 55	8	—	Higher number easier fire
Viscosity (20°C Cst)	0.7	0.72	2.3	1.3	0.78	
Firing point (°C)	464	257	338	420	—	
Flash points(°C)	17	-1 ~ 32	65.6	9 ~ 32	60	
Solidifying point (°C)	-98	-107	—	-114	-60	

- iv) Because of its low viscosity, lubrication of fuel system is required.
- v) Because methanol absorbs easily water, measures for corrosion must be considered for fuel system. The corrosion mechanism is considered to be HCOOH being produced when methanol oxidizes or reacts with water, and the HCOOH then reacts with metals having a larger ionization tendency (Al, Zn, Pb, Mg, etc.), and acrylic and urethane rubber. Thus, careful selection is necessary for materials in the fuel system.
- vi) Handling requires care due to high toxicity, (lethal dose: 30 - 100 ml. loss of eyesight: 7 - 8 ml.) though toxicity is far lower than that of petroleum. However, it must be noted that there is a possibility to drink instead of ethanol, because the small is almost same with that of ethanol.

Consideration of all these elements demonstrates that methanol possesses properties quite different from petroleum. However, it means not that methanol is impractical as a fuel for internal combustion engines, but simply that ordinary internal combustion engines are not designed to accommodate methanol. The other hand, since engines are designed to use liquid fuel, they can be easily modified to burn methanol, though it is necessary to get accurate knowledge of methanol properties.

### (3) Application of Fuel Methanol for Conventional Internal Combustion Engines

#### 1) Gasoline Engine

As the Table 8-4-1 indicates, methanol has a high octane value assuring no knocking, and allows increase of compression ratio to 10 to 12, resulting in improved SFC (specific fuel consumption). What should be noted here is that fuel supply should be doubled to offset low calorific value in volume, and carburetor should be adjusted to cover its low air-fuel ratio. Also, care is necessary to retain



corrosion resistance, which requires use of lubricants with a high alkaline value.

## 2) Gas Turbine

Since gas turbine uses continuous combustion at a constant pressure, it doesn't restrict the type of applied fuel, like a reciprocating engine. Thus in principle, virtually all types of fuel from gasoline to residual oil, including fuel methanol can be used.

Table 8-4-1 shows a comparison of jet fuel (a mixture of kerosene and gasoline) and methanol. As the Table indicates all of the subjects to be taken into consideration for use of fuel methanol are less calorific value, lower viscosity and corrosive materials. However, in case of gas turbine, above subjects are simply resolved by minor changes of fuel supply system. Therefore, methanol can be used in gas turbine much more easily than in gasoline engine and diesel engine.

As is clear in the explanation above, methanol is most easily used in conventional gasoline engine and gas turbine, though rather difficult in diesel engine. Therefore, development attention of the world is being focused upon the use of methanol in diesel engine because of features of the least fuel consumption and best thermal efficiency. In the next section we discuss the use of methanol in the diesel engine. (Data obtained from tests conducted in Japan.)

## 3) Diesel Engine

As is already explained, methanol is low in its cetane value, which makes ignition difficult in ordinary diesel engines. The following will be able to serve as possible solutions. (Table 8-4-2)

### i) Blending method

A blending method is to blend methanol and diesel oil in the fuel tank. It is injected with only one of pump and nozzle. It means that an ordinary diesel engine can meet the purpose. A problem is that methanol, by the nature, doesn't mix well with diesel oil. Thus, an appropriate co-solvent is required to

Table 8-4-2 How to Introduce Methanol – Low Cetane Number Fuel Into Diesel Engine

BLEND	BLEND Methanol and other fuel out side of engine
	BLEND Methanol and other fuel inside of engine (Dual fuel system)
Ignite	Spark assist method

eliminate the problem. Fig. 8-4-2 shows the liquid separation characteristics when Dodecanal ( $C_{12}H_{25}OH$ ), a co-solvent considered most appropriate for methanol, is used.

For example, when 10% Dodecanal is mixed into methanol and diesel oil mixture, methanol solution into diesel oil increases from 0% to 20% at a liquid temperature of 35°C.

Fig. 8-4-3 shows the methanol-gas oil engine performance compared to conventional operation by gas oil. The injection timings were fixed to give the best brake specific heat consumption (BSHC) at the brake mean effective pressure (BMEP) of 0.5 Mpa. As long as the methanol mixing ratio is less than 40 vol%, BSHC does not change very much.

Smoke density, on the other hand, is reduced significantly when blending methanol. A reduction in  $NO_x$  concentration can also be seen, but the engine noise increases about one dB with the use of the methanol blends.

Fig. 8-4-4 shows the effect of methanol on the engine performance, at BMEP of 0.72 MPa. Both engine noise and  $NO_x$  concentration change slowly with the increase in methanol. The smoke density decreases significantly with increasing methanol content. BSHC becomes the smallest at a methanol content of 20 vol%. The improvement in BSHC is stronger in the heavier load range.

ii) Dual fuel system

Since with this system, mixture is made inside the engine cylinder, a set of fuel tanks, injection pump and injector are required for methanol and diesel oil, respectively. In this case, injection of diesel oil should account for 10 - 20%, and injection timing of methanol is designed to advance a bit than that of diesel oil. The engine performance resulting from the dual fuel system is almost identical to that of the

Fig.8-4-2 Solubility Ratio of Methanol in Diesel Fuel

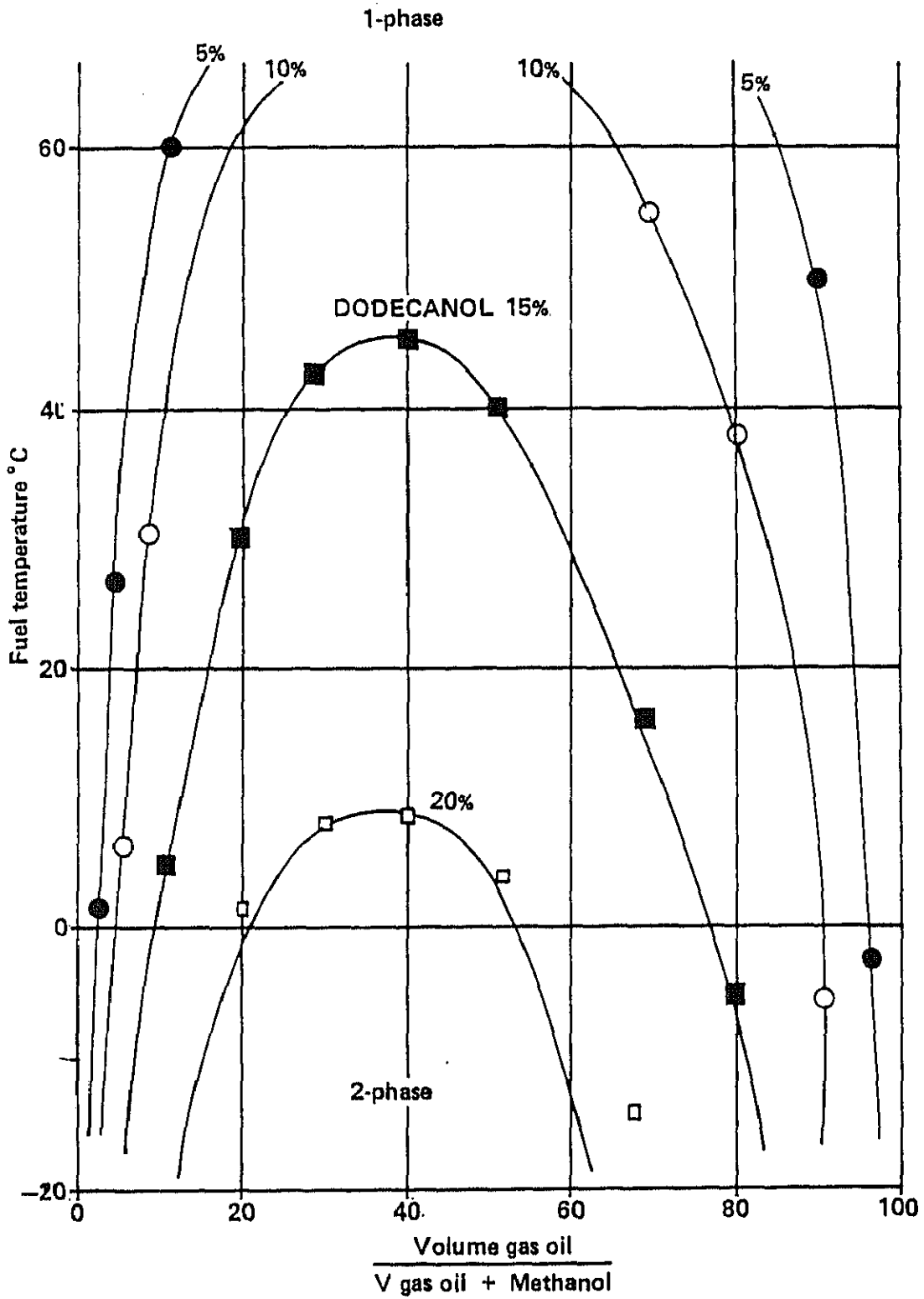


Fig.8-4-3 A Comparison of the Engine Performance with Methanol – Gas Oil Blends and with Gas Oil.

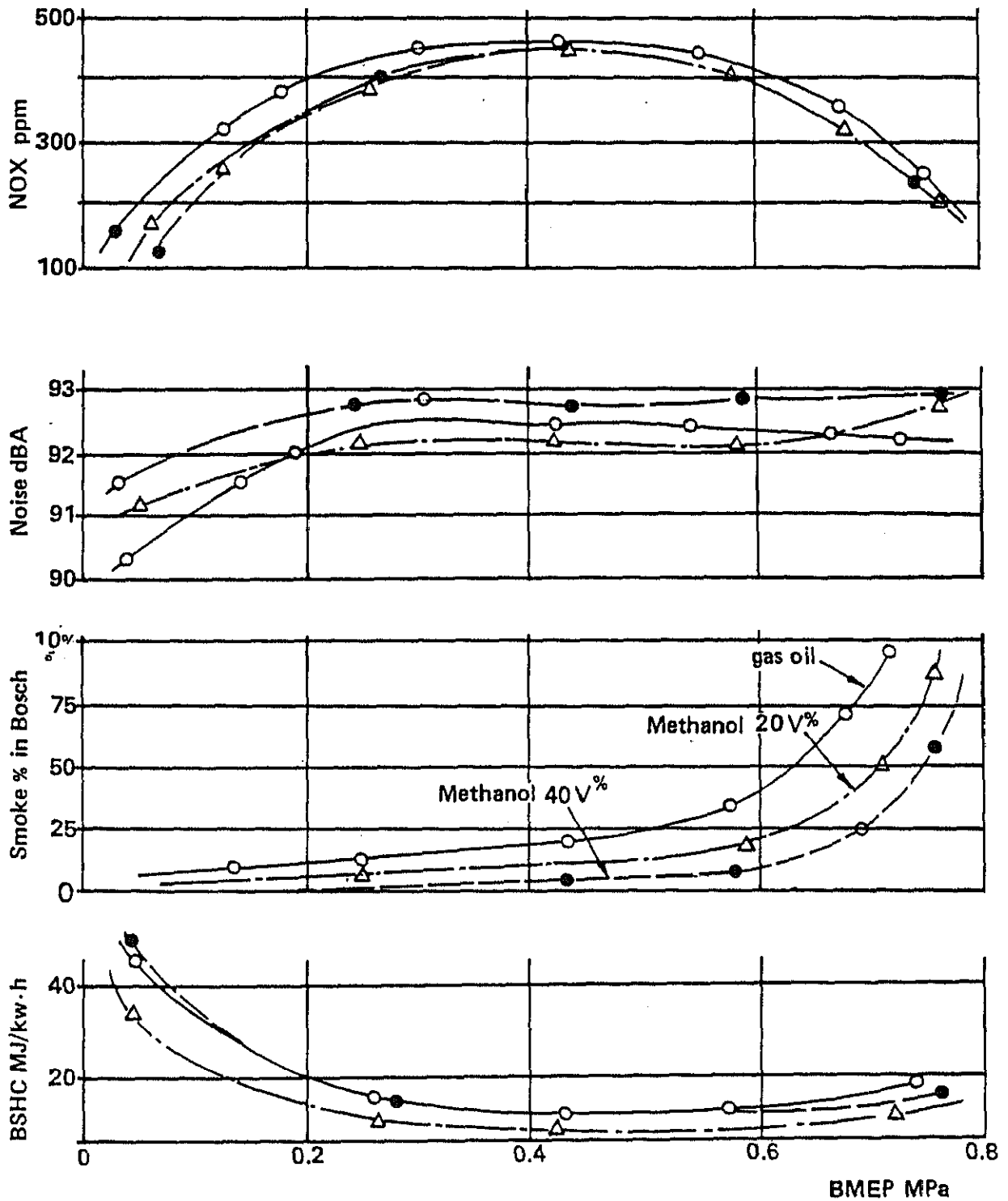


Fig. 8-4-4 Changes in Engine Performance for Different Methanol (Gas Oil) Mixing Ratios  
(BMEP = 0.72 MPa)

