

Improve benzene production from refinery sources

A new cracking route reduces cost of benzene while coproducing ethylene

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Benzene is the building block for several major commodity petrochemicals. It is mainly recovered from refinery catalytic reforming streams and furthered purified for petrochemical applications. Traditional product processing steps alkylate high-purity (99.9 wt%) benzene with ethylene to form ethylbenzene (EB)—the precursor compound for styrene monomer (SM). Similarly, benzene can be alkylated with propylene to form cumene—a precursor of phenol. To achieve such high-purity levels, HPI companies use aromatic extraction, an expensive processing step.

trates the economic benefits of steam-cracking reformate to recover benzene and increase ethylene production.

Sources of benzene. The three main sources from which benzene is recovered are catalytic reforming, pyrolysis gasoline and coke-oven gas. Benzene from refinery catalytic reforming sources is about 55% of the global production, including associated toluene conversion.¹ Benzene produced from pyrolysis gasoline, as a byproduct of ethylene manufacturing, accounts for about 40% of the global production. Approximately 5% of the worldwide benzene production is generated from coal sources, such as coke-oven gas, and less than 1% from C₃/C₄ conversion. Nearly 19% of the benzene

is produced via toluene conversion processes; however, the vast majority of toluene is originated from catalytic-reforming sources.¹ In the U.S. which accounts for 30% of the global benzene market, catalytic-reforming sources represent 70% of the benzene production, including associated toluene conversion processes.¹ Pyrolysis gasoline accounts for 28% and coal sources for 2%.

Aromatics extraction is the most common method to recover and purify benzene. Reformate from catalytic reforming naphtha is typically composed of 65–82 wt% aromatics and 18–35 wt% nonaromatics; it consists essentially of C₅ to C₇ paraffins, mostly iso-paraffins. Separating benzene from C₇ iso-paraffins by conventional distillation is difficult due to the azeotrope characteristics of C₇ iso-paraffins as well as cyclohexane

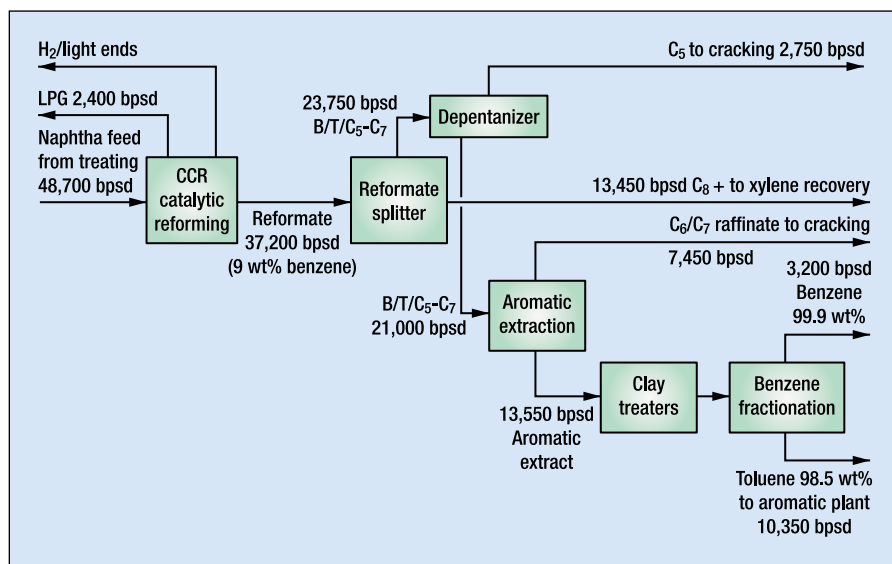


Fig. 1. Optimized conventional extraction (base case) method simplified block diagram.

and methyl-cyclopentane. Due to improved alkylation technology, lower purity benzene (about 98 wt%) can be considered for EB and cumene production. By lowering the purity requirement for these alkylation methods, benzene can be separated from product stream via new routes.* One new processing concept produces 98 wt% purified benzene, while co-producing ethylene via steam cracking at an estimated 30% lower cost. The purified benzene has adequate purity and can be used to meet the growing demand for alkylation and cyclohexane production. The presented case history illus-

and methyl-cyclopentane.

About 50% of the global benzene production, normally at 99.9 wt% purity, is used to produce EB—a precursor to styrene. An additional 23–25% of the global benzene production is used to produce cumene—a precursor to phenol. Cumene and EB are produced via alkylation of benzene with propylene and ethylene respectively. Further, about 13–15% of benzene demand is used to manufacture cyclohexane—a precursor to cyclohexanol, which is in turn a precursor for adipic acid.

Recent advances in alkylation catalysts now enable using purified benzene in the range of 90–99 wt% purity while still

* Patent application on the subject has been filed with the U.S. Patent Office.

Table 1. U.S. Gulf Coast spot market price averages in U.S./mt for benzene and naphtha as registered in month of December of the respective years.

	1990	1991	1992	1993	1994	1995	1996	1997	1998
Benzene	473	361	289	248	351	257	323	326	219
Naphtha	310	186	194	129	167	157	220	171	106
Differential	163	175	95	119	184	100	103	155	113

Table 2. Binary benzene and C₆/C₇ paraffins azeotropic characteristics

Component	Benzene, wt. %	Boiling Point, °C	Azeotropic B P, °C
Cyclohexene	85	82.1	79.5
Cyclohexane	55	81.0	77.5
Methylcyclopentane	10	72.0	71.5
n-Hexane	5	68.5	69.0
2,4 Di-methyl-pentane	48.5	80.5	75.0
2,3 Di-methyl-pentane	79.5	86.0	79.0
2,2 Di-methyl-pentane	46.5	79.0	76.0
n-Heptane	99.3	98.5	80.0
Tri-methyl-butane	50.5		76.5

meeting downstream products' specifications.² The following case study shows that the price of purified benzene, 98 wt%, is reduced by U.S.\$90/t compared with nitration-grade benzene (99.9 wt%).

TECHNOLOGY OVERVIEW

Most large-scale, high-conversion refineries have catalytic reforming units, which can provide aromatic-rich streams. In most cases, these aromatic rich streams are used as octane-enhancing blending components. However, in many other cases, they are used as a petrochemical source for downstream aromatics production. Several relevant process technologies are:

Conventional continuous catalytic reforming (CCR). Since 1980, this process was the most widely used method to reform naphtha, produce reformate and feed to downstream aromatics extraction units.³ In a conventional CCR process, the C₇+ naphtha cut is converted to 65–82 wt% aromatics and 18–35 wt% nonaromatics—mostly iso-paraffins along with 0.1–1.0 wt% olefins. The reformate is routed to aromatic extraction that recovers over 99% of the benzene at high purity while meeting all other product specifications. Full-range naphtha is pre-fractionated to C₅/C₆, 95°C cut point, to form C₇+ cut. The C₇+ cut is hydrotreated to less than 0.5 ppm sulfur along with organic nitrogen and oxygen removal. The hydrotreated C₇+ proceeds to a multi-stage reformer, operating at about 500 °C and 5 kg/cm²-g. The CCR unit produces reformate, hydrogen-rich gas and light ends. In the business-case example, 37,200-bpsd and 1.670 million tpy (1.67 MMtpy) of reformate, containing 9.0 wt% benzene, is generated from a common Middle East naphtha, which in turn is produced from 12 vol% atmospheric cut of 31°API crude oil.

Aromatic extraction. Several licensed technologies are available to recover and purify benzene by solvent extraction or extractive distillation. The solvent extraction process, is the most widely used process.⁴ Extractive distillation technologies hold a significant market share for single aromatics or either benzene/toluene or toluene/xylene extraction.^{5,6}

Reformate from catalytic reforming unit, depending on downstream considerations, is fed to a reformate splitter. Two

streams are produced—benzene/toluene cut with C₅–C₇ nonaromatic and C₈+ aromatic cut with traces of C₈ nonaromatics. The benzene/toluene cut is normally sent to depentanizer, followed by aromatic extraction process. Thus, a pure benzene/toluene mixture and a raffinate—mostly C₆/C₇ nonaromatic containing about 0.5–1.0% benzene and 1.0–2.0% toluene—is recovered.

The C₈+ cut is routed to the aromatics plant, probably for p-xylene production. The preferred outlet for the raffinate and the C₅ would be as feed to steam crackers; however, in most cases, these streams are used as gasoline blendstock. The aromatics-extract stream is treated in a clay tower to remove trace olefins and to comply with color specifications followed by benzene/toluene fractionation.

GLOBAL MARKET PRODUCT SPECIFICATION

The estimated 2001 global benzene production is about 33 MMtpy with approximately 30% of this production located in the U.S. About 50% of the global benzene supply is dedicated to EB production. Benzene specifications are based on ASTM test methods; the key tests are:

- Solidification temperature
- Relative density
- Color
- Boiling range
- Acidity and residual sulfur
- Nonaromatics content.

The solidification test is an important specification. The freezing temperature of pure benzene is +5.5°C. Contamination of benzene with C₆/C₇ or toluene will depress the freezing temperature. The freezing point specification of nitration-grade benzene, which is the most commonly available grade, is above 4.85°C. This implies that the level C₆/C₇ impurities is less than 0.5 wt%. A boiling range test of 1.0°C that includes benzene boiling temperature of 80.1°C or a density test at 15°C that indicates benzene density between 0.8820–0.8860 are less sensitive to impurities. In reality, the nitration-grade benzene is available at over 99.9 wt% with impurities of toluene and C₆/C₇. The specifications for organic nitrogen, sulfur, olefins, acetylene, moisture, etc., are highly dependent on particular downstream users.

Table 1 shows the historical spot market prices of nitration-grade benzene as well as naphtha in U.S. dollars based on U.S. Gulf Coast delivery.⁷ The global price averages for the same period were similar.

Due to new benzene-concentration restrictions in the U.S. gasoline pool—about 1.0 vol%, (more stringent in California)—the recent market trend shows a reduction in the price of benzene and increase in the price of toluene as a high-octane blending component. One method to lower benzene concentration is alkylating the reformate heart cut (dilute benzene) with ethylene from FCC offgases to form alkylated-benzene fuel.⁸ A portion of the pyrolysis gasoline, partially hydrotreated and containing about 35 wt% benzene, may no longer be acceptable as a gasoline blending stock and would be diverted to benzene recovery. This added benzene could be offset by reducing toluene hydrodealkylation processing which in 2000, accounted for 6% of the global benzene market. The net result will have a minimal impact on the potential market for purified benzene.

The global SM production rate is approximately 21.5 MMtpy and is the largest benzene consumer—about 16.5 MMtpy. The issues of impure benzene are focused on EB; however, the same logic is applicable for cumene and cyclohexane production.

Table 3A. Deheptanizer Section

	Stream-1, Kg/h	Stream-2, Kg/h	Stream-3, Kg/h	Stream-4, Kg/h	Stream-5, Kg/h
i-C ₄ H ₁₀	25	25	0	0	0
n-C ₄ H ₁₀	595	665	0	70	0
C ₅ mix	10,865	12,980	0	2,115	0
n-C ₆ H ₁₄	5,780	5,980	10	210	0
i-C ₆ H ₁₄	10,750	10,920	10	180	0
M-Cyclo C ₅	335	562	3	230	0
Cyclo C ₆	45	58	2	15	0
n-C ₇ H ₁₆	5,330	4,835	535	5	35
i-C ₇ H ₁₆	9,890	9,910	10	10	20
M-Cyclo C ₆	100	30	110	5	35
Benzene	18,060	18,595	10	495	50
Toluene	58,420	305	59,660	5	1,540
p-Xylene	10,310	2	10,383	0	75
o-xylene	13,785	2	13,878	0	95
m-xylene	22,420	5	22,570	0	155
EB	9,730	10	10,245	0	533
C ₈ Non-aromatics	0	0	35	5	30
C ₉ + Aromatics	22,510	0	22,510	0	0
Total	198,950	64,884	139,971	3,345	2,568

Table 3B. Ethylene plant cracking section

	Stream-6, Kg/h	Stream-7, Kg/h	Stream-8, Kg/h	Stream-9, Kg/h
Hydrogen	0	0	6,260	0
CO	0	0	155	0
Methane	455	0	25,040	0
Acetylene	0	0	2,140	0
Ethylene	0	0	98,560	0
Ethane	83,335	42,005	42,720	0
MAPD	0	0	595	0
Propylene	0	55	15,055	0
Propane	41,270	3,040	3,095	0
C ₄ mix	455	0	8,790	0
C ₅ mix	0	0	2,035	0
C ₆ NA	0	0	835	0
C ₇ NA	0	0	250	0
C ₈ NA	0	0	30	0
Benzene	0	0	22,620	0
Toluene	0	0	1,550	0
Xylene + EB + Sty	0	0	850	0
Heavy fuel oil	0	0	0	4,690
Total	125,515	45,100	230,580	4,690

uid or vapor phase.²¹ Over 90% of cyclohexane is converted to cyclohexanol, which is a precursor to adipic acid. Purified benzene of 98 wt% purity is likely to meet all specifications to manufacture cyclohexanol and adipic acid.

To illustrate the advantages of the new process arrangement, a conventional benzene extraction method will be used as a benchmark for comparison purposes. In Fig. 1, reformat is fed to a reformat splitter—4.75-m ID, 70 trays, 28 MM Kcal/hr reboiler using 40 kg/cm²-g steam—followed by a depentanizer, 2.5-m ID 32 trays 5.5 MM Kcal/hr reboiler using 3.5 kg/cm²-g steam. Depentanizer bottoms consisting of benzene/toluene/C₆/C₇ proceeds to aromatic extraction unit, using 52 tph steam at 40 kg/cm²-g.

Aromatic extract—benzene/toluene product followed by clay treating—is routed to benzene/toluene fractionation, 2.75-m ID 65 trays and 9.6 MM Kcal/hr reboiler using 8.0 kg/cm²-g steam.

Reformat-splitter bottoms composed of C₈+ with traces of C₈ nonaromatic are sent to the aromatics plant. The RON (research octane number) of raffinate is about 55–60; thus, its value as a steam-cracker feed is at least \$20–24/t higher than its value as gasoline blending stock. The cost of transport raffinate from the Middle East to European or Japanese steam crackers is about \$20–24/t. This price differential represents a credit advantage of approximately \$7.0–8.4 million/yr for the business case. This easily justifies the added investment to the steam cracker to handle the mixed-feed design, particularly when considering the incremental propylene and BTX production.

The raffinate—C₆/C₇ product, about 7,450 bpsd along with 2,750 bpsd C₅ mix—provides additional feed to the steam cracker, thus raising production 144 Mtpy of ethylene, 67 Mtpy of propylene, 34 Mtpy of C₄ mix and 18 Mtpy of benzene.

BUSINESS CASE

A possible business case is based on gas feed (C₂/C₃) to a steam cracker in the Arabian Gulf. Assume that the contemplated ethylene plant is adjacent to a large petroleum-refining complex that includes a catalytic reforming unit. High-severity catalytic reforming technology is assumed as a benzene source.³ Note that this concept as shown is also applicable to liquid crackers, which are common in Europe and Asia-Pacific.

Definition. The primary feed to the steam cracker is ethane/propane mix with C₂/C₃ = 2.0 by weight. However addi-

tional olefins, about 17% of total ethylene, are produced by cracking C₅-C₇ produced from light reformat. The design capacity of the ethylene plant is 850 Mtpy based on 350 days per year on stream operation. C₄-mix product is exported to battery limits. About 147 Mtpy of contained ethylene is diverted to produce 535 Mtpy of EB, which is converted to 500 Mtpy of SM. The balance of the ethylene, 703 Mtpy, is directed toward polymer-grade ethylene production.

Pyrolysis gasoline, containing about 79 wt% benzene, is hydrotreated and fractionated to form 189 Mtpy benzene with 98 wt% purity, meeting organic nitrogen, sulfur, moisture and olefins specifications. The toluene could be a feed to a disproportionation unit to produce 195 Mtpy of pure benzene and 290 Mtpy xylenes, in addition to the pre-existing 395 Mtpy of xylenes, 85 Mtpy EB and 215 Mtpy of C₉+ aromatics. Additional 20 Mtpy toluene and 5 Mtpy benzene as byproducts from styrene production, will make the complex self sufficient in benzene totaling about 400 Mtpy.

The business case as described is based on using 12-mol% dilute ethylene from demethanization section of a front-end deethanizer ethylene plant at 27.5 kg/cm²-g as a feed to the alkylation unit.^{22,23} The plant design includes a front-end acetylene hydrogenation unit. Feasibility studies have shown that this dilute-ethylene scheme for alkylation applications is far more economical when compared with 72-mol% ethylene feed that could be recovered from the C₂ fractionator, or a pure ethylene as produced.

Process description. Fig. 2 is a simplified diagram that illustrates the technical concept of the business case. The total material balance (Fig. 2) is based on 8,400 hours/yr operation and is shown in Tables 3A–3E.

About 2 MMtpy pre-cut naphtha, (48,700 bpsd), feeds the CCR unit to produce 1.67 MMtpy reformat, (37,200 bpsd, Stream -1) along with 75 Mtpy (2,400 bpsd) C₃/C₄ and hydrogen-rich gas, about 87 mol%.²⁴ The reformat, along with 1,200 bpsd from benzene purification, is charged to deheptanizer V-101. This is a 75 tray, 4-m ID column operating at 1.5 kg/cm² at the top. The deheptanizer bottom is re-boiled by steam at 8.0 kg/cm²-g, with a re-boiling duty of 19.5 MM Kcal/hr. Toluene-xylene-rich bottom product proceeds to the toluene column—4-m ID, 55 trays and re-boiling duty of 17.5 MM Kcal/hr using steam at 40 kg/cm²-g.

Table 3C. Product fractionation

	Stream-10, Kg/h	Stream-11, Kg/h	Stream-12, Kg/h	Stream-13, Kg/h	Stream-14, Kg/h
Hydrogen	0	0	0	6,260	6,260
CO	0	0	0	155	155
Methane	5	0	0	25,040	25,040
Ethylene	83,500	0	0	17,530	500
Ethane	115	5	0	850	850
Propylene	15	15,015	25	0	0
Propane	0	55	0	0	0
C ₄ mix	0	5	8,790	0	0
C ₅ mix	0	0	90	0	0
Benzene	0	0	0	0	50
C ₆ /C ₇	0	0	0	0	15
Total	83,630	15,070	8,905	49,835	32,870

Table 3D. Product hydrotreating

	Stream-15, Kg/h	Stream-16, Kg/h	Stream-17, Kg/h	Stream-18, Kg/h
Hydrogen	0	0	142	0
C ₄ mix	15	51	0	70
C ₅ mix	910	1,130	0	2,115
C ₆ mix NA	623	210	0	870
C ₇ mix NA	232	15	0	260
C ₈ mix NA	29.7	0.3	0	30
Benzene	20,230	2,390	0	22,620
Toluene	1,530	20	0	1,550
Xy/EB/Styrene	849.3	0.7	0	858
Total	24,419	3,817	142	28,373

Table 3E. Benzene fractionation

	Stream-19/4, Kg/h	Stream-20, Kg/h	Stream-21/5, Kg/h	Stream-22, Kg/h
n-C ₄ H ₁₀	70	0	0	0
C ₅ saturated	2,115	0	0	0
M-Cyclo C ₅	230	65	0	65
Cyclo C ₆	15	125	0	125
i-C ₆	180	5	0	5
n-C ₆	210	45	0	45
i-C ₇	10	155	20	135
n-C ₇	5	100	35	65
C ₇ Napht	5	45	35	10
C ₈ NA	0	30	30	0
Benzene	495	22,125	50	22,075
Toluene	5	1,545	1,540	5
C ₈ aromatic	0	858	858	0
Total	3,340	25,098	2,568	22,530

Condensing duty up to 65°C is provided by tempered-water cycle, loading the heat on seawater at 33°C and releasing at 38°C. Overhead flow, Stream (2), is directed to a steam cracker for ethylene and propylene production. The bottom flow of toluene and heavy aromatic is directed to aromatic plant for xylene and additional benzene production. The benzene-rich, light-reformate feed along with C₂/C₃ feed produce the desired nominal 850 Mtpy ethylene, 126 Mtpy propylene and 70 Mtpy C₄-mix products, containing 50 wt% butadiene.

The contemplated ethylene plant will require eight or nine steam crackers, including a spare, consisting of two liquid crackers and six to seven gas crackers. All furnaces include steam generation, at 120 kg/cm²-g at 510°C. The liquid crackers require 15.0 MM Kcal/hr (18%) more firing duty than raffinate cracking due to the high-benzene feed, producing approximately 11 tph additional steam at 120 kg/cm²-g at 510°C, and about 9.0 tph steam at 8.0 Kg/cm²-g. However, cracking may require 9.0 tph of additional dilution steam at

Table 4. Relative capital investment for conventional benzene extraction vs. purified benzene method, total installed cost (TIC) basis

Description	Base case (Conventional extraction method), U.S.\$MM	Business case (Purified benzene method), U.S.\$MM
Catalytic reforming 50,000 bpsd, about \$130 MM investment	Base	Base
Depentanizer	2.5	–
Reformate splitter	4.5	–
Deheptanizer	–	4.0
Aromatic extraction 93 Mtpy (21,000 bpsd) 71 wt% (64 vol%) benzene/toluene	42	–
• Extraction solvent initial charge/inventory	2.0	–
Clay treating, 67 Mtpy, (2 x 50% units)	3.0	–
Dehexanizer	–	2.5
Benzene column	3.5	2.0
Toluene Column	–	3.5
Six or seven gas crackers (Capital investment of about \$140 MM)	Base	Base
Two liquid crackers with steam generation 8.5-tph ethylene each at BL	30	34.5
Quench oil/quench water/dilution steam	Base	1.0
Cracked-gas compressor	Base	0.5
Pyrolysis gasoline stripper and hydrotreating (High benzene adjustment)	Base	1.0
Ethylbenzene plant (Capital investment about \$32 MM)	Base	Base
Total	87.5	49.0

Table 5. Utilities consumption figures (estimated)

Description	Base case	Business (new) case
Depentanizer • Steam @3.5 kg/cm ² – g	11 tph	–
Reformate splitter • Steam @40 kg/cm ² – g	65 tph	–
Deheptanizer Steam @8.0 kg/cm ² – g	–	40 tph
Aromatic Extraction • Steam @40 kg/cm ² – g • Power	52 tph 700 kw	– –
Clay Treating • Steam @8.0 kg/cm ² – g	2.5 tph (based on 5% of aromatic extraction unit steam consumption)	–
Dehexanizer • Steam @3.5 kg/cm ² – g	–	12.5 tph
Benzene column • Steam @8.0 kg/cm ² – g	20 tph	9.5 tph
Toluene Column • Steam @40 kg/cm ² – g	–	43 tph
Liquid Crackers (2) • Steam @120 kg/cm ² – g/510°C • Fuel Gas LHV	– 83.0 MM Kcal/hr	–11 tph (export) 98.0 MM Kcal/hr
Quench oil/quench water/ dilution steam • Steam @8.0 kg/cm ² – g	–	9 tph –9 tph (export)
Cracked Gas Compressor • Steam @120 kg/cm ² – g/510°C	Base	1.8 tph (500 Kw)
Heat rejection (air coolers/ tempered water/sea water) • Power	1,450 kw	–

8.0 Kg/cm²-g over conventional raffinate and also slight increase in 120 kg/cm²-g 510°C motive steam to drive the cracked-gas compressor. The liquid crackers are equipped with quench oil system. Cracked gas from the quench oil is merged with cracked gas from the gas crackers and enters quench water system.

Cracked gas from quench water system is cooled to 42°C, water and heavy liquids are separated, prior to entering a five-stage cracked-gas compressor driven by a 38,000 kW steam turbine. The suction pressure is about 0.4 kg/cm²-g and discharge pressure at the after-cooler KO drum, is 32 kg/cm²-g. Traces of H₂S and CO₂ are removed by inter-stage caustic wash.

The bulk of the benzene and C₅+ liquids are condensed in the inter-stage, after-stage and pre-dryer coolers, separated in the drums and then routed to olefins saturation. The cracked gas is chilled further to 15°C before drying. The dry gas is reheated to 60°C prior to entering acetylene reactor. The acetylene-free dry cracked gas is chilled before proceeding to de-ethanizer. The de-ethanizer overhead, at about 29.5 kg/cm²-g and consisting of CH₄, C₂H₄, C₂H₆ and traces of propylene, is chilled and routed to the “sloppy cut” demethanizer. The demethanizer overhead at 27.5 kg/cm²-g consists of all produced hydrogen and methane along with about 17.5% of the ethylene, about 2% of ethane and less than 1 ppm propylene. After cold recovery, this stream becomes the dilute ethylene feed at 26.0 kg/cm²-g and 35°C to the ethylbenzene plant, Stream 13.

The bottom of the de-ethanizer proceeds to the depropanizer. Overhead from the depropanizer proceeds to propylene fractionation, and the bottom is sent for C₄-mix recovery. The C₅-pyrolysis gasoline from C₄ mix recovery is combined with pyrolysis gasoline from compressor drums and sent to a second-stage hydrotreating unit. The first stage saturates di-olefins and the second stage saturates olefins and reduces sulfur content to below 0.5 ppm.

Hydrotreated pyrolysis gasoline, about 79 wt% benzene, is routed to benzene purification, which consists of two columns. The first column, V-102, has 60 trays, 2-m ID dehexanizer with 6.5 MM Kcal re-boiling duty with 3.5 kg/cm²-g steam. The dehexanizer overhead, containing the bulk of C₆ and some residual benzene, is recycled to deheptanizer V-101. Bottom product proceeds to benzene column V-103. The benzene column is 40 trays, 2-m ID and 5.0 MM Kcal re-boiling duty with 8 kg/cm²-g steam. Overhead product is 98 wt% benzene, and the bottom is toluene-rich stream that is recycled to deheptanizer V-101. The purified benzene (Stream 22) along with supplemental benzene from aromatic plant, react with dilute ethylene to form ethylbenzene. Vent gas from dilute ethylene production is sent to a PSA unit for hydrogen recovery and fuel gas separation. The EB is then routed to a conventional or POSM styrene plant.

Material balance. Tables 3A–3E are material balances for the described case in Kg/hr. The simulations of reformate and

Table 6. Economic Comparison Between Base Case and Business Case

Description	Base case (Conventional extraction method)	Business case (Purified benzene method)
Capital charges @ 28%	\$23.94 MM	13.72 MM
• Extraction solvent initial inventory @ 20%	\$0.40 MM	0.0
Electric power @ \$0.050/kwh	\$1.070 MM	Base
Fuel gas LHV @ \$5.0/MMKcal	\$3.49 MM	\$4.16 MM
Steam, 90% condensate recovery		
• 120 Kg/cm ² g 510°C @14/t	Base	-\$1.082 MM
• 40 Kg/cm ² g sat. @ 10.5/t	\$10.05 MM	\$3.8 MM
• 8.0 Kg/cm ² g sat. @ 8.0/t	\$1.52 MM	\$3.33 MM
• 3.5 Kg/cm ² g sat. @ 6.0/t	\$0.53MM	\$0.63 MM
Extraction solvent make up	\$0.21 MM	0.0
Total	\$41.21 MM	\$24.56 MM

benzene fractionation use SRK thermodynamic package. The streams as listed correspond to the numbers indicated in Fig. 2, which includes up to 0.5 wt% olefins.

Capital investment considerations. Table 4 represents the relative capital investments for the optimized conventional design (base case) compared to the purified benzene production design (business/new case).²⁶ The average OSBL in this comparison is 40% of ISBL. All figures are based on 2001, U.S. Gulf Coast location, curve type estimate in U.S. \$million.²⁷

Table 4 shows that the capital investment requirements for the business (new) case, based on purified benzene production method, is U.S.\$ 38.5 million less than the capital required for



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Omar J. Ghalayini is an international project development executive with over 25 years of experience with petroleum and hydrocarbon processing technologies, process-plant design, and engineering and construction. His focus is primarily on technology commercialization, project sales and international negotiations and contracting. Mr. Ghalayini held a leading project and business development role on several "first of kind" projects, including the first commercial "chemicals from coal" facilities built in the U.S., the first dedicated HGU based on partial oxidation and Rectisol technologies and the first integrated ethane recovery, nitrogen rejection and helium production plant built in North America. Mr. Ghalayini has extensive experience in developing projects in the Middle East, Indian sub-continent and China. Prior to launching Projexx Consulting International, Inc. in 1999, Mr. Ghalayini spent 15 years with Linde AG (Lotepro Corp.) as director, market development and five years with Howe-Baker Engineers, Inc., as vice president, international sales. He holds a BS degree in chemical engineering from University of Houston (ojg@projexx.com).



a conventional benzene extraction method.

Utilities. Table 5 represents approximate utilities consumption figures for base case and business (new) case. Plant water, instrument air and other minor utility consumption figures are considered negligible and are not accounted for.

Economic analysis. Table 6 is an economic comparison between the base case (conventional extraction method) and the business/new case (purified benzene method) for 185 Mtpy contained benzene. From Table 6, the business case, based on the purified benzene method, has approximately an annual U.S.\$17 million cost advantage over the conventional extraction method. This is equivalent to cost savings of more than \$90/t of contained benzene. Adjusting the economic analysis to a Middle Eastern location may increase the savings to over \$100/t of benzene.

Options. Process improvements in alkylation indicate that using purified benzene at 98 wt% in lieu of nitration grade is technologically feasible and economically advantageous. The total cost savings derived from producing benzene from refinery sources with co-production of olefins are substantial and were calculated at over \$90/t of benzene in this case study.

The potential consumption of purified benzene could account for up to 90% of the benzene market. Production of pure benzene from toluene conversion, or C₃/C₄ conversion, about 20%, could eliminate the need for aromatic extraction. The U.S. Gulf Coast is home to 92% of ethylene production—about 25 MMtpy—along with 100% of styrene and cumene. This production independently "coexists" with 1.5 MMbpsd catalytic reforming capacity.²⁶ In the rest of the world, the vast majority of steam crackers, alkylation facilities and catalytic reforming units are all located in the same proximity of other industrial complexes. In these situations, geographic synergism offers strategic advantages that would provide the added commercial gain to consider implementation of the purified benzene production concept. ■

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- Union Carbide at Texas City, Texas, and Chevron at Cedar Bayou, Texas, have operated DMF acetylene recovery units.
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- Communication with ABB Lummus.
- University of Ghent, Belgium; Kellogg Brown & Root, Houston, Texas; Linde AG, Munich, Germany, and ABB Lummus.
- Chevron-Phillips Chemical and ExxonMobil are the largest cyclohexane producers.
- U.S. Patent 5,880,320, "Combination process for manufacturing ethylene ethylbenzene and styrene," D. Netzer. Patent applications are pending in India, China and Europe.
- U.S. Patent 6,177,600. 23, "Combination process for manufacturing ethylene benzene and alkylated benzene," D. Netzer.
- The depleted naphtha 200°F cut point is about 460 Mtpy and the C₃/C₄ 75 Mtpy could be used as a feed to the cracker and produce 200 Mtpy ethylene.
- Based on communication with Stone & Webster.
- This is based on Corpus Christi, Texas, to Pascagoula, Mississippi, section of Gulf Coast. Total U.S. reforming capacity is 3.5 MMbpsd, and global capacity 11MMbpsd.
- Accuracy of +25%/-15%.