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The industrial application of MIP-DCR technology in catalytic cracking units

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Abstract: To reduce dry gas and coke yields and further optimize product distribution, an industrial trial of the MIP-DCR technology was conducted on a 2.10 Mt/a fluid catalytic cracking (FCC) unit of a Sinopec subsidiary. The results show that under conditions of identical catalyst and generally comparable overall operating parameters. At the same time, the Conradson carbon residue (CCR) of the feedstock for MIP-DCR was markedly higher than that for the conventional MIP process. Compared with the MIP operation case, the dry gas yield of MIP-DCR decreased by 0.04 percentage points; the yields of liquefied petroleum gas (LPG) and gasoline increased by 0.16 and 0.26 percentage points, respectively. The total yield of LPG, gasoline and light diesel oil rose by 0.04 percentage points, and the total yield of target liquid products remained roughly unchanged. The slurry oil yield decreased by 1.00 percentage point, whereas the coke yield increased by 1.00 percentage point. The above results basically reflect the technical characteristics of MIP-DCR in lowering dry gas yield, promoting heavy oil conversion, and optimizing product distribution, demonstrating promising potential for industrial application. Nevertheless, the impacts of increased coke yield and the difference in feedstock Conradson carbon residue on overall economic benefits still require comprehensive evaluation combined with long-term operational data.

Keywords: product distribution; dry gas; liquefied gas; stable gasoline; diesel; coke.

INTRODUCTION

Fluid catalytic cracking (FCC) is one of the most important secondary conversion processes in the petroleum refining industry, undertaking the core task of converting heavy oil into high-value light products.[1,2] With the increasing trend of crude oil heavying and deterioration, as well as continuous tightening of clean

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fuel specifications, conventional FCC processes are confronted with increasingly severe challenges in optimizing product distribution. Against this backdrop, the Research Institute of Petroleum Processing, Sinopec (hereinafter referred to as RIPP), has developed a novel FCC technology designated MIP-DCR, aiming to reduce dry gas and coke yields, based on the basis of the Maximizing Iso-Paraffins (MIP) process and the MIP-CGP clean gasoline technology for maximizing iso-paraffins and propylene production.

The core concept of this technology originates from an in-depth understanding of FCC reaction chemistry. According to the Haag–Dessau dual-reaction mechanism, cracking of alkanes over acidic catalysts proceeds via both monomolecular protolytic cracking and bimolecular carbenium-ion cracking.[3-5] In the FCC process, dry gas formation is closely associated with the above reaction pathways together with thermal cracking pathways, among which monomolecular protolytic cracking and thermal cracking serve as major contributors to dry gas generation.[6] Protolytic cracking preferentially occurs at the initial stage of feedstock-catalyst contact and is favored by high temperature and low hydrocarbon partial pressure; thermal cracking reactions exhibit stronger temperature sensitivity. Therefore, lowering the initial contact temperature between regenerated catalyst and feedstock and optimizing the initial contact conditions can suppress non-selective cracking pathways such as protolytic cracking and thermal cracking at the early reaction stage, so as to reduce the yields of dry gas and coke.[7]

Technical implementation of the MIP-DCR process relies mainly on two key engineering measures. First, the conventional pre-lift section at the riser bottom is retrofitted into a catalyst pre-lift mixer, in which high-temperature regenerated catalyst from the regenerator is thoroughly mixed with low-temperature regenerated catalyst cooled by an external heat exchanger. This markedly reduces the temperature of the mixed catalyst prior to its contact with feedstock. Second, the feedstock preheat temperature is moderately raised within the allowable range of overall reaction heat balance to improve feedstock atomization, enlarge catalyst-oil contact area, and optimize the initial catalyst-oil contact conditions.[8] These two measures function synergistically: on one hand, local overheating at the instant of catalyst-oil contact is alleviated to inhibit non-selective reactions including protolytic cracking and thermal cracking; on the other hand, feedstock atomization and dispersion are enhanced to mitigate the formation of thermal coke. This paper mainly presents the industrial application results of MIP-DCR technology on a 2.10 Mt/a FCC unit at a Sinopec subsidiary.

EXPERIMENTAL

Device overview

The fluid catalytic cracking (FCC) unit of a Sinopec subsidiary has a designed capacity of 2.10 Mt/a. Its reaction section originally adopted the MIP process developed by the Research Institute of Petroleum Processing, Sinopec (RIPP). In contrast, the regeneration section

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employed the proprietary series main-air regeneration technology with fast bed-turbulent bed developed by Luoyang Petrochemical Engineering Corporation, Sinopec (LPEC). The product slate was designed to produce gasoline with high octane number and low olefin content. To further reduce dry gas and coke yields and increase total liquid yield, the RIPP-developed MIP-DCR process was adopted for the reaction section. On the basis of the original fast bed-turbulent bed series main-air regeneration technology, oxygen-enriched regeneration was applied in the regeneration section to intensify coke burning and reduce the carbon content on regenerated catalyst. After revamping, the product slate was adjusted to produce high-octane, low-olefin gasoline, together with liquefied petroleum gas (LPG) and light diesel oil. The main revamping works of the unit include installing a pre-lift mixer at the bottom of the riser reactor and adding a new downward-flow finned-tube external heat exchanger. High-temperature regenerated catalyst from the regenerator is thoroughly mixed with low-temperature regenerated catalyst cooled by the external heat exchanger in the pre-lift mixer, so as to decrease the temperature of the mixed regenerated catalyst and thereby lower the initial contact temperature between feedstock and catalyst.

Feedstock oil

The properties of feedstocks used in the two fluid catalytic cracking processes are compared in Table I. It can be seen from Table I that the MIP-DCR feedstock is generally similar to the MIP feedstock, while exhibiting characteristics of slightly lower density, lighter distillation range, higher saturation content, and marginally lower sulfur content and higher Conradson carbon residue (CCR). Specifically, the density of the MIP-DCR feedstock is 933.1 kg/m^3, which is 2.61 kg/m^3 lower than that of the MIP feedstock; its sulfur content is 2.12%, representing a decrease of 0.05 percentage points relative to the MIP feedstock. In terms of group composition, the saturated content of the MIP-DCR feedstock increases by 4.03 percentage points compared with the MIP feedstock, whereas aromatic, resin, and asphaltene contents decrease by 3.26, 0.70, and 0.07 percentage points, respectively. This indicates that the contents of unsaturated components such as aromatics, as well as heavy components including resins and asphaltenes, are reduced in the MIP-DCR feedstock. Regarding the distillation range, the 10%, 30%, 50%, 70%, 90%, and 95% distillation temperatures of the MIP-DCR feedstock are all lower than those of the MIP feedstock. Among them, the 10% distillation temperature decreases by 23.8 °C, demonstrating that its overall fraction is lighter. It is noteworthy that the CCR of the MIP-DCR feedstock reaches 1.61%, an increase of 0.81 percentage points compared with the value of 0.80% for the MIP feedstock. This constitutes the primary deterioration difference between the two feedstocks, which may raise the coke formation tendency and regeneration load during FCC processing. Overall, the MIP-DCR feedstock is slightly superior to the MIP feedstock in density, sulfur content, and group composition, but possesses markedly higher CCR. Therefore, a comprehensive evaluation combining product distribution and regeneration capacity should be performed when comparing the performance of the two processes.

Process operation parameters

The main operating parameters of the two fluid catalytic cracking processes are listed in Table II. It can be observed from Table II that, compared with the MIP operation case, the MIP-DCR operation generally features higher throughput, elevated reaction temperature, increased feedstock preheat temperature, and a slight rise in disengager pressure. Specifically, the fresh feed throughput under the MIP-DCR condition is 193 t/h, which is higher than 184 t/h for the MIP case; the disengager pressure increases from 0.343 MPa (absolute) to 0.350 MPa

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(absolute), representing an increment of 0.007 MPa. In terms of reaction temperature, the outlet temperature of the first reaction zone rises from 532 °C to 534 °C, and the temperature of the second reaction zone increases from 525 °C to 529 °C, corresponding to increments of 2 °C and 4 °C, respectively. Under MIP-DCR operation, the temperature of the mixed catalyst inside the pre-lift mixer is controlled at 679 °C, and the feedstock preheat temperature is 8 °C higher than that in the MIP case. In addition, quench injection was not adopted in either operating case. Overall, the MIP-DCR process runs at a higher throughput, higher reaction temperature, higher feedstock preheat temperature, and marginally higher disengager pressure. The reaction conditions are intensified relative to the MIP case, which may enhance the cracking reaction intensity. Nevertheless, MIP-DCR reduces the initial catalyst–oil contact temperature via mixing hot and cold regenerated catalysts. Accordingly, its influence on conversion, product distribution, coke formation, and pressure balance needs to be comprehensively evaluated on the basis of product distribution and industrial operational data.

TABLE I. Properties of the feedstock

Item	MIP	MIP-DCR	Difference
Density (20 °C), kg/m^3	935.71	933.1	-2.61
Sulfur content, wt%	2.17	2.12	-0.05
Carbon residue, wt%	0.8	1.61	0.81
Group composition, wt%			
Saturates	37.67	41.7	4.03
Aromatics	60.66	57.4	-3.26
Resins	1.30	0.6	-0.70
Asphaltenes	0.37	0.3	-0.07
Distillation range, °C			
10%	355.8	332	-23.80
30%	401.1	391	-10.10
50%	437.7	432	-5.70
70%	478.2	475	-3.20
90%	535.8	533	-2.80
95%	558.6	556	-2.60

TABLE II. Main operating parameters

Item	MIP	MIP-DCR
Disengager pressure, MPa (abs)	0.343	0.350
Fresh feed rate, t/h	184	193
Outlet temperature of the first reaction zone, °C	532	534
Temperature of the second reaction zone, °C	525	529
Feed preheat temperature, °C	252	260
Pre-lift mixer temperature, °C	—	679
Terminating agent (water) injection rate, t/h	0	0

Catalyst

The identical catalyst was used for both fluid catalytic cracking processes, and the properties of the regenerated catalyst under different operating conditions are compared in Table III. As shown in Table III, when listed in the order of MIP-DCR/MIP, the contents of metallic

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contaminants and specific surface area of regenerated catalyst for the two cases are basically comparable. The contents of iron, nickel and vanadium are approximately 1991/1991, 1182/1190 and 1712/1715 mg/kg, respectively, and the specific surface areas are 146/147 m^2/g. These results indicate no obvious difference in metal deposition extent and surface properties between the two catalysts. The carbon content on the regenerated catalyst for MIP-DCR is 0.04 wt%, slightly higher than the value of 0.01 wt% for MIP, yet still at a low level. Its microactivity is 57.0, 0.6 units lower than 57.6 for MIP, showing a slight decrease in catalytic activity. In contrast, the difference in particle size distribution is relatively remarkable. The fraction of 0–20 μm fines in regenerated catalyst for MIP-DCR decreases substantially from 1.8% under MIP to 0.02%, and the fraction of 20–40 μm particles drops from 14.8% to 13.18%. Meanwhile, the fractions of 40–80 μm and >80 μm particles rise from 47.4% and 36% to 49.71% and 37.09%, respectively. Overall, the regenerated catalyst under MIP-DCR exhibits evidently coarser particle size and substantially reduced fines. This change is beneficial for reducing catalyst fines loss; however, insufficient fines may deteriorate fluidization quality and catalyst circulation. The corresponding impacts on catalyst transportation and cyclone separation system should also be monitored. In summary, with the same catalyst employed, the regenerated catalysts under the two operating cases show minor differences in heavy metal contamination, specific surface area, and microactivity, and their contribution to the discrepancy in product distribution is relatively limited.

TABLE III. Properties of the regenerated catalyst

Item	Regenerated catalyst	
	MIP	MIP-DCR
Specific surface area, m^2/g	146	147
Carbon content, wt%	0.01	0.04
Fe, mg/kg	1991.6	1990.8
Ni, mg/kg	1181.7	1189.6
V, mg/kg	1712	1715
Microactivity	57.6	57
Particle size distribution (φ), %		
0–20 μm	1.8	0.02
20–40 μm	14.8	13.18
40–80 μm	47.4	49.71
>80 μm	36	37.09

RESULTS AND DISCUSSION

Product distribution

The product distributions of fluid catalytic cracking before and after the implementation of the MIP-DCR process are compared in Table IV. It can be seen from Table IV that the total yield of target liquid products (defined as the sum of light oil yield and liquefied petroleum gas yield) is roughly comparable for the two operating cases, which are 78.54% and 78.60% for MIP and MIP-DCR, respectively, corresponding to a marginal increase of only 0.06 percentage points for the latter. In terms of product composition, under the MIP-DCR operation case, gasoline yield rises by 0.27 percentage points, and liquefied petroleum gas (LPG) yield increases by 0.17 percentage points relative to the MIP case, while light

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diesel oil yield decreases by 0.38 percentage points, and dry gas yield drops slightly by 0.04 percentage points. This reveals that under elevated reaction temperature and intensified heavy oil conversion, part of the components in the diesel fraction are converted into gasoline and LPG. Meanwhile, dry gas yield is still slightly reduced even with heavier feedstock and more severe reaction conditions, demonstrating that the MIP-DCR process can suppress non-selective reactions such as protolytic cracking and thermal cracking to a certain extent. With respect to the conversion of heavy components, slurry oil yield decreases from 10.04% to 9.04%, a reduction of 1.00 percentage point, which indicates substantially enhanced heavy oil cracking depth. Coke yield increases from 7.2% to 8.20%, an increment of 1.00 percentage point, which is equivalent to the reduction in slurry oil yield. Considering that the feedstock Conradson carbon residue (CCR) rises from 0.80% to 1.61% under MIP-DCR, accompanied by increases in both reaction temperature and feedstock preheat temperature, the elevated coke yield is mainly attributed to the enhanced coke-forming tendency of the feedstock and deeper conversion extent. Overall, MIP-DCR improves heavy oil conversion capacity and markedly reduces slurry oil yield, whereas the total yield of target liquid products does not increase correspondingly. Given the significantly higher CCR of MIP-DCR feedstock, the comparable magnitude of slurry oil reduction and coke increment suggests that the improved conversion depth of heavy components is achieved mainly at the expense of increased coke formation.

TABLE IV. Product distribution

Item	MIP	MIP-DCR	Difference
wt%			
Dry gas	4.07	4.03	-0.04
LPG[a]	16.20	16.36	0.16
Gasoline	40.83	41.09	0.26
Light diesel oil	21.64	21.26	-0.38
Slurry oil	10.05	9.05	-1
Coke	7.21	8.21	1
Total	100	100	
Light oil yield + LPG	78.67	78.71	0.04

[a] liquefied petroleum gas

Product quality

The properties of dry gas from fluid catalytic cracking before and after the implementation of the MIP-DCR process are compared in Table V. It can be seen from Table V that, relative to the MIP operation case, dry gas obtained under the MIP-DCR operation case exhibits a reduced volumetric fraction of hydrogen, increased hydrocarbon components, and elevated weighted average molecular weight, indicating a compositional transition of dry gas from relatively “hydrogen-

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rich" to "hydrocarbon-rich". This variation is associated with the increases in reaction temperature, feedstock preheat temperature, and throughput of MIP-DCR, which is built upon the serial dual-reaction-zone framework of the MIP technology, as well as the intensified heavy oil cracking, secondary conversion, and hydrogen transfer reactions.[9] Specifically, the volumetric fraction of hydrogen decreases from 24.47% to 20.30%. Hydrocarbon components including methane, ethane, ethylene, propane and propylene all rise to varying degrees, and the weighted average molecular weight increases from 18.2 g/mol to 18.9 g/mol, suggesting an improvement in the calorific value and light hydrocarbon recovery value of dry gas. The reduction in hydrogen fraction is partly attributed to the enhanced hydrogen transfer reactions: olefins accept active hydrogen donated by hydrogen donors to form paraffins, suppressing the tendency for olefins to undergo further cracking, dehydrogenation, and condensation to produce hydrogen. On the other hand, the higher saturated content and lower aromatic content in the MIP-DCR feedstock also reduce the sources of hydrogen generation via dehydrogenation to some extent. The increased hydrocarbon components reflect deeper cracking severity, with more C_2 and C_3 species entering the dry gas stream. The volumetric fraction of H_2S rises from 3.79% to 4.15%, which originates from the increased cleavage of sulfur-containing compounds with enhanced conversion depth. Combined with the decreased H_2S content in liquefied petroleum gas (LPG) shown in Table VI, sulfur redistribution likely occurs between gaseous and liquid products, with partial sulfur migrating into dry gas. Overall, after implementing the MIP-DCR process, the proportion of hydrogen in dry gas decreases while hydrocarbons, especially ethylene and propylene, increase. However, the elevated H_2S concentration requires intensified dry gas desulfurization to satisfy the requirements of downstream processing and environmental compliance.

The properties of liquefied petroleum gas (LPG) from fluid catalytic cracking before and after the implementation of the MIP-DCR process are compared in Table VI. It can be observed from Table VI that, relative to the MIP operation case, the LPG obtained under the MIP-DCR operation generally shows a trend of propylene enrichment and reduced C_4 olefins. This phenomenon essentially represents the combined effect of elevated reaction severity and synergistically enhanced hydrogen transfer, secondary cracking, and isomerization reactions on the distribution of C_3/C_4 fractions. Specifically, the volumetric fraction of propylene increases from 32.03% to 38.69%, accompanied by simultaneous rises in propane and isobutane. Except for a slight increase in isobutene, C_4 olefins including trans-2-butene, cis-2-butene, and 1-butene decrease markedly, indicating an improved recovery value of propylene in LPG and a reduction in the total amount of C_4 olefins. The above variations are closely related to the further intensified conversion of heavy components in the MIP-DCR process: higher reaction temperature and conversion depth promote the cracking of large-

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molecular hydrocarbons to form propylene, while a portion of C_4 olefins undergoes further cracking, hydrogen transfer, or isomerization.[10] The volumetric fraction of H_2S decreases from 9.14% to 4.41%, which may be attributed to the altered conversion pathway of sulfur-containing species under MIP-DCR operating conditions, where more sulfur is transferred into dry gas in the form of H_2S. This change is beneficial for the refining and desulfurization of LPG, yet it should be considered together with the increased H_2S concentration in dry gas. Combined with the decreased slurry oil yield and increased coke yield shown in the product distribution results, it can be concluded that the MIP-DCR process improves LPG composition toward higher propylene content and lower H_2S by increasing conversion depth.

TABLE V. Properties of the dry gas

Item	MIP	MIP-DCR
φ, vol%		
Hydrogen	24.47	20.30
Hydrogen sulfide	3.79	4.15
Methane	28.27	31.60
Ethane	13.44	14.33
Ethylene	10.65	12.41
Propane	0.03	0.25
Propylene	0.21	0.98
$\leq C_2$	52.36	58.34
$\geq C_3$	0.89	1.72
Oxygen	0.60	0.41
Nitrogen	14.33	11.71
Carbon monoxide	0.86	0.67
Carbon dioxide	2.70	2.70
Weight-average molecular weight	18.20	18.90

TABLE VI. Properties of the LPG

Item	MIP	MIP-DCR
φ, vol%		
Propane	9.50	11.90
Propylene	32.03	38.69
Isobutane	16.44	17.50
n-Butane	3.99	3.83
trans-2-Butene	8.62	6.37
1-Butene	6.53	6.37
Isobutene	7.48	7.59
cis-2-Butene	6.09	3.27
$\leq C_2$	0.15	0
$\geq C_5$	0.03	0.07
Hydrogen sulfide	9.14	4.41

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The properties of stabilized gasoline from fluid catalytic cracking before and after the implementation of the MIP-DCR process are compared in Table VII. It can be seen from Table VII that, compared with the MIP operation case, the stabilized gasoline under the MIP-DCR operation generally exhibits the following characteristics: reduced olefin content, increased octane number, lighter distillation fraction, together with elevated sulfur content and vapor pressure. The forward shift of gasoline distillation range, slight reduction in density, and rise in vapor pressure indicate an enhanced degree of lightening. The volumetric fraction of olefins decreases to 25.2%, while the volumetric fractions of aromatics and benzene increase to 27.4% and 0.69%, respectively, and the research octane number (RON) rises to 92.9. These results demonstrate that part of the olefins in gasoline are converted into high-octane components such as aromatics and iso-paraffins via hydrogen transfer, aromatization, alkylation, and isomerization reactions. Consequently, the octane number increases rather than decreases while olefin content is reduced, which is consistent with the technical feature of MIP-series technologies for “olefin reduction with octane preservation”. The increased sulfur content may be associated with feed sulfur distribution, deeper conversion extent, and redistribution of sulfur-containing species within the gasoline fraction. Combined with the previously observed trend of higher H_2S in dry gas and lower H_2S in LPG, it can be inferred that the transformation and distribution pathways of sulfur among gaseous and liquid products have been modified. The induction period decreases from 913 min to 905 min, with only an 8 min reduction, and remains at a relatively high level. This indicates that although increased sulfur content and vapor pressure may impose adverse impacts on service performance, the reduced olefin content helps improve oxidation stability, and the overall stability of stabilized gasoline remains favorable. In summary, the MIP-DCR process modifies stabilized gasoline toward lower olefin content and higher octane number; nevertheless, the increases in vapor pressure and sulfur content require corresponding optimization of stabilizer operation, desulfurization, and blending management.

The properties of fluid catalytic cracking (FCC) diesel before and after the application of the MIP-DCR process are compared in Table VIII. Compared with the MIP operation case, the overall quality of diesel deteriorated under the MIP-DCR condition, mainly manifested as increased density, elevated sulfur content, and reduced cetane index. This phenomenon is attributed to the enhanced reaction severity and intensified secondary conversion in the MIP-DCR process. Crackable paraffins in the diesel fraction are preferentially converted, while refractory low-cetane components such as polycyclic aromatic hydrocarbons and cycloaromatics are relatively enriched, shifting the diesel composition toward higher aromaticity and lower cetane quality. In terms of distillation and physical properties, the initial boiling point and flash point rise to 211.5 ℃ and 103.5 ℃, respectively, indicating

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that more light components are cracked or stripped off. The 50%, 90%, and 95% distillation temperatures decrease slightly, while the density increases significantly. This variation demonstrates that the diesel fraction does not undergo simple lightening; instead, the forward shift of the distillation range is accompanied by a density increment caused by enhanced aromaticity. The cetane index decreases from 18.6 to 17.7, further confirming the deterioration of ignition performance. The sulfur content increases by 0.15 percentage points from 2.84% to 2.99%, which is associated with the deepened conversion degree, redistributed and enriched sulfur-containing species in the diesel fraction, as well as the heavier feedstock, higher Conradson carbon residue, and altered sulfur distribution. Combined with the reduced slurry oil yield and increased coke yield in product distribution, it can be concluded that the MIP-DCR process improves heavy oil conversion at the cost of deteriorated diesel quality to a certain extent. In practical industrial operation, intensified hydrodesulfurization, hydrodenitrogenation, and aromatic saturation of diesel are required. Moreover, a comprehensive trade-off between conversion depth and diesel quality should be implemented.[11]

TABLE VII. Properties of the stabilized gasoline

Item	MIP	MIP-DCR
Density (20 °C), kg/m^3	738.9	736.5
Vapor pressure, kPa	48.3	63
Research octane number (RON)	91.45	92.9
Induction period, min	913	905
Sulfur content, %	0.125	0.161
Olefins (φ), vol%	26.9	25.2
Benzene (φ), vol%	0.39	0.69
Aromatics (φ), vol%	25.38	27.4
Distillation range, °C		
IBP[a]	36.7	30.5
10%	52.2	45.5
50%	100.8	95
90%	175.5	174
FBP[b]	200.6	202.5

[a] initial boiling point; [b] final boiling point

The properties of fluid catalytic cracking slurry oil before and after the implementation of the MIP-DCR process are compared in Table IX. It can be observed from Table IX that, compared with the MIP operation case, the slurry oil density increases from 1139 kg/m^3 to 1152.5 kg/m^3, and the solids content rises from 1.77% to 2.6% under the MIP-DCR operation case. The increased slurry oil density, on the one hand, reflects the relative enrichment of heavy components such as aromatics, resins, and asphaltenes, accompanied by enhanced aromaticity and a higher degree of aromaticity further reduced crackability. On the other hand, the elevated solids content also contributes partially to the rise in slurry oil density.

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The higher solids content indicates the enrichment of solid particles including entrained catalyst fines, catalyst loss and coke powder in slurry oil. This will aggravate the abrasion of equipment such as slurry oil pumps, heat exchangers and nozzles, raise the risk of coking and blockage at the fractionator bottom and within the slurry oil system, and impair the subsequent utilization of purged slurry oil. This phenomenon arises from the intensified secondary cracking, hydrogen transfer, and heavy aromatic conversion of the MIP-DCR process, which reduce the slurry oil yield. With a similar purge strategy, the purge flow rate decreases, leading to the relative enrichment and concentration of refractory polycyclic aromatic hydrocarbons, resins, asphaltenes, as well as solid matter like catalyst fines and coke powder in the remaining slurry oil. Combined with the decreased slurry oil yield and increased coke yield in the product distribution results, it can be concluded that the MIP-DCR process deepens conversion extent while deteriorating slurry oil properties and worsening the operating conditions of the slurry oil system. Therefore, under MIP-DCR operating conditions, emphasis should be placed on monitoring slurry oil density, solids content, and purge flow rate, optimizing the purge ratio and solids removal measures, so as to guarantee the long-term stable operation of the slurry oil system while improving heavy oil conversion capacity.

TABLE VIII. Properties of the diesel

Item	MIP	MIP-DCR
Density (20 °C), kg/m^3	959.2	991.2
Flash point (closed cup), °C	98.8	103.5
Cetane index	18.6	17.7
Sulfur content, wt%	2.84	2.99
Distillation range, °C		
IBP[a]	203.9	211.5
10%	237	238.5
50%	278.5	275.5
90%	344.9	342.3
95%	355.1	353

[a] initial boiling point

TABLE IX. Properties of the slurry oil

Item	MIP	MIP-DCR
Density (20 °C), kg/m^3	1139	1152.5
Solid content	1.77	2.6

CONCLUSION

This paper presents the industrial application results of the MIP-DCR technology on a 2.10 Mt/a fluid catalytic cracking unit at a Sinopec subsidiary. The results show that under conditions of identical catalyst and generally comparable overall operating parameters but remarkable differences in feedstock properties,

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compared with the MIP operation case, the dry gas yield under the MIP-DCR operation case decreases by 0.04 percentage points; the yields of liquefied petroleum gas (LPG) and gasoline increase by 0.16 and 0.26 percentage points, respectively. The total yield of LPG, gasoline, and light diesel oil rises by 0.04 percentage points, and the total yield of target liquid products remains roughly unchanged. Meanwhile, the slurry oil yield decreases by 1.00 percentage point, whereas the coke yield increases by 1.00 percentage point. To a certain extent, the above results reflect the technical characteristics of MIP-DCR in lowering dry gas yield, promoting heavy oil conversion, and improving product distribution, demonstrating promising potential for industrial application. Nevertheless, considering that the Conradson carbon residue (CCR) of the MIP-DCR feedstock is markedly higher than that of the MIP feedstock, together with the increased coke yield, its impacts on overall economic benefits still require comprehensive evaluation incorporating long-term operational data and feedstock property differences.

DECLARATION OF AI USE

During the preparation of this manuscript, the authors used ChatGPT (GPT-4, OpenAI) to assist with English language translation and polishing to improve readability. The authors have carefully reviewed, edited, and verified all AI-assisted output, and take full responsibility for the accuracy, originality, and final content of this publication.

И З В О Д

ИНДУСТРИЈСКА ПРИМЕНА MIP-DCR ТЕХНОЛОГИЈЕ У ПОСТРОЈЕЊИМА ЗА КАТАЛИТИЧКО КРЕКОВАЊЕ

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У циљу смањења приноса сувог гаса и кокса и даље оптимизације расподеле производа, спроведено је индустријско испитивање MIP-DCR технологије на постројењу за каталитичко флуидизовано крековање (FCC), капацитета 2,10 милиона тона годишње, у оквиру једне од компанија корпорације *Sinopec*. Резултати показују да су, при употреби истог катализатора и генерално упоредивим укупним радним параметрима, вредности Конрадсоновог остатка угљеника (CCR) у сировини за MIP-DCR процес биле значајно више него у случају конвенционалног MIP процеса. У поређењу са MIP процесом, принос сувог гаса у MIP-DCR процесу смањен је за 0,04 процентна поена, док су приноси течног нафтног гаса (LPG) и бензина повећани за 0,16, односно 0,26 процентних поена. Укупан принос LPG-а, бензина и лаког гасног уља повећан је за 0,04 процентна поена, док је укупан принос циљних течних производа остао приближно непромењен. Принос суспензионог уља смањен је за 1,00 процентни поен, док је принос кокса повећан за 1,00 процентни поен. Наведени резултати у основи одражавају техничке карактеристике MIP-DCR технологије, које се

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огледају у смањењу приноса сувог гаса, интензивирању конверзије тешких фракција нафте и оптимизацији расподеле производа, што указује на њен значајан потенцијал за индустријску примену. Ипак, утицај повећаног приноса кокса, као и разлике у вредностима Конрадсоновог остатка угљеника у сировинама, на укупне економске ефекте процеса захтева свеобухватну процену засновану и на дугорочним подацима о раду постројења.

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REFERENCES

1. S. M. Sadrameli, *Fuel* **173** (2016) 285 (https://doi.org/10.1016/j.fuel.2016.01.047)
2. A. C. Akah, E. Al-Shafei, Q. Xu, M. AlHerz, Z. S. Qureshi, M. A. Bari Siddiqui, A. Aitani, *Chem. Eng. J. Adv.* **23** (2025) 100794 (https://doi.org/10.1016/j.ceja.2025.100794)
3. Y. F. Zuo, L. Ye, W.J. Yang, B. Peng, J. Zhang, X. T. Shu, Y. H. Xu, J. C. Liu, *ACS Catal.* **15** (2025) 1576 (https://doi.org/10.1021/acscatal.4c05323)
4. J. H. Zhang, H. H. Shan, X. B. Chen, C. Y. Li, C. H. Yang, *Ind. Eng. Chem. Res.* **52** (2013) 6366 (https://doi.org/10.1021/ie4005496)
5. G. Caeiro, R. H. Carvalho, X. Wang, M.A.N.D.A. Lemos, F. Lemos, M. Guisnet, F. Ramôa Ribeiro, *J. Mol. Catal. A: Chem.* **255** (2006) 131 (https://doi.org/10.1016/j.molcata.2006.03.068)
6. A. F. H. Wielers, M. Vaarkamp, M. F. M. Post, *J. Catal.* **127** (1991) 51 (https://doi.org/10.1016/0021-9517(91)90208-l)
7. J. H. Gong, Y. H. Xu, Z. Cai, K. Q. Xie, *Pet. Process. Petrochem.* **44** (2013) 6 (http://www.sylzyhg.com/EN/Y2013/V44/I6/6)
8. L. G. Qiao, *Pet. Refin. Eng.* **45** (2015) 7 (https://doi.org/10.3969/j.issn.1002-106X.2015.06.002)
9. C. X. Han, Y. H. Xu, B. Lu, W. Wang, *Chem. Ing. Tech.* **95** (2023) 97 (https://doi.org/10.1002/cite.202200120)
10. W. J. Yang, Y. H. Xu, X. T. Shu, X. Wang, X. H. Bai, Y. F. Zuo, Y. Ouyang, *Appl. Energy* **349** (2023) 121665 (https://doi.org/10.1016/j.apenergy.2023.121665)
11. J. S. Gao, C. M. Xu, *Pet. Sci. Technol.* **17** (1999) 471 (https://doi.org/10.1080/10916469908949729)

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